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EVALUATION OF COAL GASIFICATION/COMBINED CYCLE POWER PLANT FEASIBILITY

AT THE

SEWELLS POINT NAVAL COMPLEX

NORFOLK, VIRGINIA

DEPARTMENT OF THE NAVY
ATLANTIC DIVISION
NAVAL FACILITIES ENGINEERING COMMAND
NORFOLK, VIRGINIA

FINAL REPORT

JULY 1981

POPE, EVANS AND ROBBINS INCORPORATED
CONSULTING ENGINEERS

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20. ABSTRACT (Continue on reverse side if necessary and identify by block number)

This study evaluates the feasibility of installing a coal gasification/combined cycle cogeneration plant at Sewells Point Naval Complex, Norfolk, Virginia. The study addresses current gasification technology, combined cycle thermodynamics, environmental control requirements, and conventional coal fired cogeneration cycles. The utility interface, site considerations and economic analyses are also presented. The study concludes that a coal gasification/combined cycle cogeneration plant supplying 50 MW of electric power and 290,000 lb/hr of steam is technically feasible.

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1.0 INTRODUCTION

The United States Navy has long been aware of the need to seek alternatives to the use of relatively high cost fuel oil and natural gas for generating steam and power for its shore establishment. Directions that such efforts have taken include conversion to coal as the primary fuel (see References 1-6) and extensive conservation efforts with significant emphasis on cogeneration (see References 3, 7 and 8). The purpose of this Energy Showcase Project, funded by NAVFAC and the Department of Energy, is to determine a most suitable type of cogeneration system for implementation and installation at the Sewells Point Naval Complex (SPNC), Norfolk, Virginia.

A cogeneration facility, in general terms, may be defined as one which produces, from one fuel, electric energy and steam or other forms of thermal energy which are used for heating or cooling purposes. Thus these plants simultaneously produce two forms of useful energy: electricity and heat. When properly designed, they use less fuel than would be needed to produce the power and heat separately. Factors that must be considered in evaluating cogeneration plants include electric and steam demands and their coincidence; space requirements; pollution control; labor for operating and maintenance; reliability; and capital, operating and labor costs. The foregoing factors include those that are generic and those that are site specific. The determination of a suitable optimum system requires a careful evaluation of all these factors coupled to a life cycle cost analysis.

Cogeneration systems generally take two forms: selective energy or total energy. In the former, the cogeneration plant operates in parallel with the utility and provides only part of the power requirements and perhaps only part of the thermal energy. In the total energy system, the cogeneration plant provides all of the energy requirements of the

facility and is independent of the local utility. The energy production in either case may be derived either through a topping cycle facility in which the fuel energy input to the plant is first used to produce useful power output and the rejected heat from power production is then used to provide the thermal energy, or through a bottoming-cycle facility in which the fuel energy input to the system is first applied to a useful thermal energy process with the reject heat energy therefrom used for power production. Combinations of these cycles are also possible.

The focus for SPNC is on coal gasification/ combined cycle power plants. For comparison purposes, we also address a conventional coal-fired electric and steam power plant.

This report contains the variety of elements needed to make this assessment. In Section 2.0, we provide a complete overview of coal gasification technology focusing our attention on currently available technology. Section 3.0 presents the study of combined cycle thermodynamics using loads representative of those at Sewells Point. In Section 4.0, environmental controls for the gasification/combined cycle are discussed. Section 5.0 presents the conventional coal-fired electric and steam power plant. The cooperation and interest of the local utility, Virginia Electric Power Company (VEPCO), is vital to schemes such as those under consideration here; results from interviews with VEPCO are in Section 6.0. Section 7.0 sets forth site considerations. In Section 8.0 will be found a life cycle cost analysis and a life cycle energy benefit analysis for each of the preferred alternative cogeneration candidates. Note that we provide separately bound Executive Summary and Appendices for this report.

The remainder of this section provides a status report on the facility, an analysis of the existing systems, a detailed discussion of the loads, both current and projected, and a brief look at coal availability.

1.1 SPNC Facility Status

The central power plant for SPNC is located in Building P1, Exhibit 1-1. At present, steam is generated by seven oil fired boilers and is used to generate some electricity (for peak shaving purposes) but primarily to provide steam services for pier cold iron, base industrial processes and building heating.

A summary of the condition, rating and firing capability of each boiler in P1 is shown in Table 1-1. There are 7 boilers in service at the present time. The eighth boiler, capable of firing pulverized coal and No. 6 oil, has been installed but not commissioned as yet. Aside from this new boiler the facility is old by industrial standards.

Boilers designated 55, 56 and 57 are oil-fired (No. 6 oil). These boilers are 75,000 lb/hr capacity each and cannot be converted to coal.

Boilers designated 59, 60, 61 and 62 are of somewhat larger capacity. The first three have 100,000 lb/hr capacity; No. 62 has a firing rate of 115,000 lb/hr. These 4 boilers burn No. 6 oil at present, but are capable of firing coal as well. There is a plan to retube the high pressure boilers in the near future, and possibly the air preheaters as well.

Vanadium deposit on boiler tubes has been a persistent problem in the plant. Uncontrolled emission of V_2O_5 has

SITE PLAN FOR EXISTING POWER PLANT
SEWELLS POINT NAVAL COMPLEX

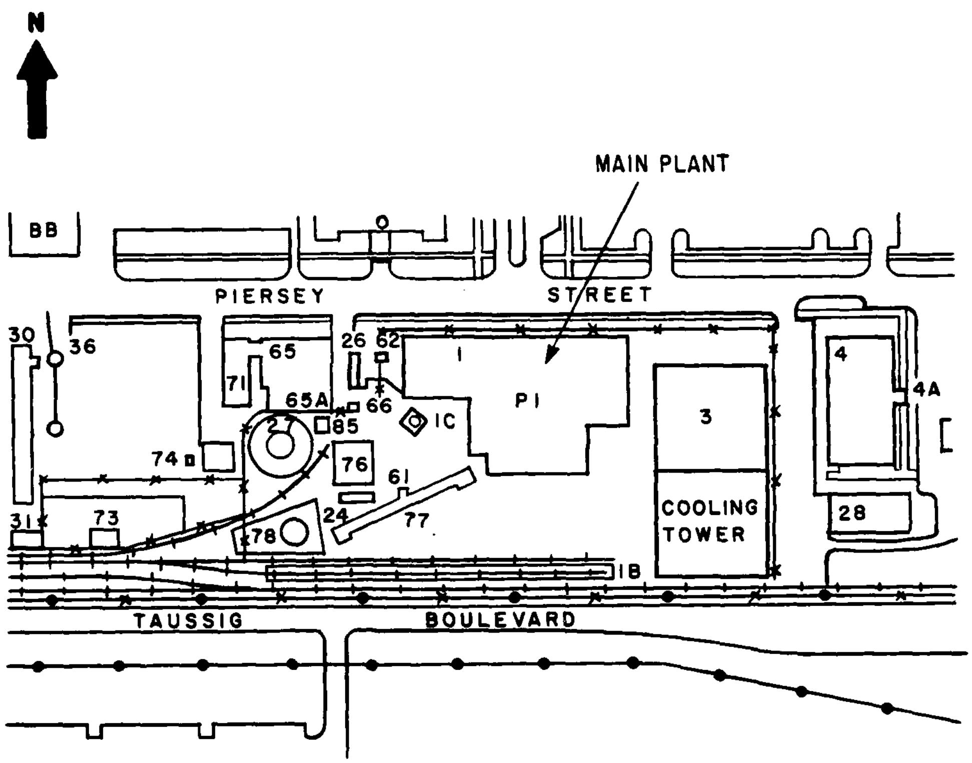


EXHIBIT I-1

TABLE 1-1

SEWELLS POINT NAVAL COMPLEX
EXISTING BOILERS
BUILDING P-1

Boiler Number	Manufacturer	Year Manufactured	Rating (lb/hr)	Pressure (psig)	Fuel	Remarks
55	Riley	1940	75,000	135	Oil	Packaged, oil-fired, No. 6, boiler.
56	Riley	1940	75,000	340	Oil	Packaged, oil-fired, No. 6, boiler.
57	Riley	1940	75,000	340	Oil	Packaged, oil-fired, No. 6, boiler.
59	Combustion Engineering	1942	100,000	410	Oil/Coal	Originally fired pulverized coal.
60	Combustion Engineering	1942	100,000	410	Oil/Coal	Originally fired pulverized coal.
61	Combustion Engineering	1942	100,000	410	Oil/Coal	Originally fired pulverized coal.
62	Combustion Engineering	1944	115,000	410	Oil/Coal	Oriignally fired pulverized coal.
-	Riley	1980	200,000	340	Oil/Coal	New boiler, not operational.

also been causing ecological problems. However, plant management is planning to feed MgO in the near future to control this phenomenon arising from burning No. 6 fuel oil.

When the eighth boiler comes on line, total capacity at P1 will be 840,000 lb/hr. Firm capacity with the largest boiler out of service will be the same as the current total capacity: 640,000 lb/hr.

Another boiler plant is located near the waterfront. It consumes the waste products of the activity and produces steam from the heat generated by incineration. Two other existing plants are used as peaking units during winter. Data relevant to these other facilities is in Table 1-2. While these plants are not directly involved in the potential coal conversion/coal gasification, their capacities and loads are important for a total perspective of the SPNC facilities. Thus in the load management to be discussed later, this total capacity of 410,000 lb/hr will be part of the system outputs and demands.

Since four of the boilers in P1 and the new one are capable of burning coal and since much of the original coal handling equipment has been retained and maintained, a project for SPNC, P-985, has been developed to reconvert those boilers to pulverized coal firing. While we will offer a considerable discussion of the project in later sections of this report, in this facility status discussion report it is pertinent to describe it briefly here.*

---

*It should be noted that as of July 1981 this project was deprogrammed. This effect on the economics of the cycles considered here will be seen later in Section 8.0.

TABLE 1-2

SEWELLS POINT NAVAL COMPLEX
SATELLITE BOILER PLANTS

BREEZIE POINT - SP85

Boiler Manufacturer	Year Manufactured	Ratings (lbs/hr)	Pressure (psig)	Fuel	Remarks
Riley	1942	75,000	125	Oil #2	Boilers originally designed to burn coal. However, coal handling equipment is no longer there
Riley	1942	75,000	125	Oil #2	

SOUTH ANNEX-NH 200

Boiler Manufacturer	Year Manufactured	Ratings (lb/hr)	Pressure (psig)	Fuel	Remarks
Babcock and Wilcox	1917	60,000	125	Oil #6	Only 3 years life expectancy.
Wix Boiler Co.		60,000	125	Oil	Mobil unit replaces a 60,000 lb/hr Babcock & Wilcox boiler which is out of commission.
Cleaver Brooks		11,000	125	Oil	Supplementary Unit.
Vapor		3 @ 3200	125	Oil	Supplementary Units.

REFUSE PLANT

Boiler Manufacturer	Year Manufactured	Ratings (lb/hr)	Pressure (psig)	Fuel	Remarks
Foster Wheeler	1965	60,000	125	Refuse & Oil	Generally burns refuse and oil combined.
Foster Wheeler	1965	60,000	125	Refuse & Oil	

This project will construct new coal storage silos to provide the minimum capacity to supply Steam Plant P1 requirements for about 30 days or 20% of the total fuel consumption for a year, whichever is greater, in compliance with criteria of DM-3. Replacement of railway delivery trackage and coal unloading equipment will be included. In addition, the ash disposal handling system will be reworked and/or replaced as required. Flue gas particulate controls are included. Boilers will be modified as necessary and coal processing equipment will be replaced. Also boiler stacks will be raised to eliminate local fumigation problems. Inactive coal bin will be included. The existing coal storage and rail delivery system are to be demolished. Cooling towers will replace the existing spray pond.

To insure environmental compliance, new flue gas particulate controls will be added to the boilers. To handle coals currently available, new coal pulverizers will be provided. Burners will be replaced and boiler breeching will be reworked. New stacks are required. The existing coal delivery and storage system will be totally replaced as well as the ash removal equipment.

Plant P1 has limited power generation equipment. There are two 4000 kW turbogenerators in the plant which are strictly used for shaving peak demands from VEPCO.

Turbine No. 1 is under overhaul at this time. Out of 34 stages, it has already lost 18 due to component failure. When returned to service with so many missing expansion stages, the machine will be usable for 1500 kW at the most. The plant management is thinking of replacing this rotor with a new one that Allis Chalmers has promised to fabricate for them.

Turbine No. 6 is now in operation, generating around 2500/2700 kW only. This is considered adequate by the plant personnel for the purpose of peak shaving. This turbine has all the expansion stages intact.

Condenser tube leakage has been experienced in the past, but there is no record to establish the mechanics of failure. Failed tubes have never been subjected to metallurgical analysis. However, there is no steam/condensate cycle conditioning and it is conceivable that corrosion by CO_2, which is very aggressive in presence of oxygen, might have taken place. Corrosion by electrolysis was mentioned in passing, and differential aeration due to living organisms adhering to tube surfaces is a distinct possibility if there is algae in the cooling water. A positive residual of chlorine is ensured at all times in the cooling water to prevent this.

1.2 Loads At SPNC

It should be emphasized that SPNC is the single largest U.S. Navy energy consumer in the continental United States. As a consequence, opportunities for economies of scale will be present here which may not be duplicated in any other Naval facility. Indeed, as we will see, the overall steam and electric requirements are equivalent to those of a small utility.

To effectively establish complex requirements for the cogeneration schemes, loads were projected to the 1988 time period. This is the assumed date for any new system -- coal gasification/combined cycle or high pressure boiler with steam turbine -- to go on line.

The steam and electric loads at SPNC were analyzed to determine their patterns, magnitudes and special characteristics. After considerations of expected growths and federally mandated energy conservation measures, the results were projected to the design year of the proposed cogeneration project. This yielded load duration curves, monthly loads, and typical daily loads for the design year. Growth rate estimates were based upon information provided by the planning department, on data extracted from the Master Plan for SPNC and those shown in References 6 and 9.

The elements used for the projections to the 1988 base year are:

<u>For Steam Demand</u>

- Current annual steam generation is approximately 3.6×10^9 lb/year (based on FY 1979 data).

- Accounting for in-plant steam use for feedwater heating and auxiliaries and for desuperheating results in a steam export of 80% of steam generated; therefore, steam exported in 1979 is roughly 2.9×10^9 lb/year.

- A growth rate in steam requirements of 4% per year is assumed in accordance with historical data at SPNC (see Reference 9), with the Master Plan, and with the projected increases in homeporting activities.

- Energy conservation is to be implemented in concert with Executive Order 12003, 20 July 1977, requiring a 20% reduction in energy use in existing government facilities by FY 1985 measured from the base year of 1975, some of which has already been accomplished.

For Electrical Demand

- Current annual electric consumption is approximately 500×10^6 kWh/year (based on FY 1979 data).

- Historic growth rates have been close to 8% per year (see Reference 9); for purposes of this study we assume that this has been reduced to 4.5% in accord with Master Plan analyses.

- Executive Order 12003 likewise applies to the electric demand.

From these elements, we determine that total steam to be generated in 1988 will approximate 4.75×10^9 lbs/year, with export at 3.8×10^9 lbs/year. Electric requirements will rise to 600×10^6 kWh/year.

Analyses of UCAR and daily load data coupled to the projections provide the steam and electrical load duration curves; these are shown in Exhibits 1-2 and 1-3, respectively. These curves may be used for purposes of preliminary system selection and cost analyses. Aside from peak and minimum requirements shown on the curves, it is useful to define from them base and intermediate loads. The base load is usually taken as the load surpassed during 5000 hours while an intermediate load is surpassed during 2000 hours. We should emphasize that these load duration curves, taken together, are generally not useful in correlating coincidence of steam and electric demands. This is due to the possible time shift between load requirements: for example, steam demand at night and electric demand during daytime.

On Exhibit 1-2 steam exported from the main plant, the salvage plant and the peaking plants are separately identified. The contribution from the various plants is as follows:

EXPORT STEAM LOAD DURATION CURVE

PROJECTED TO 1988

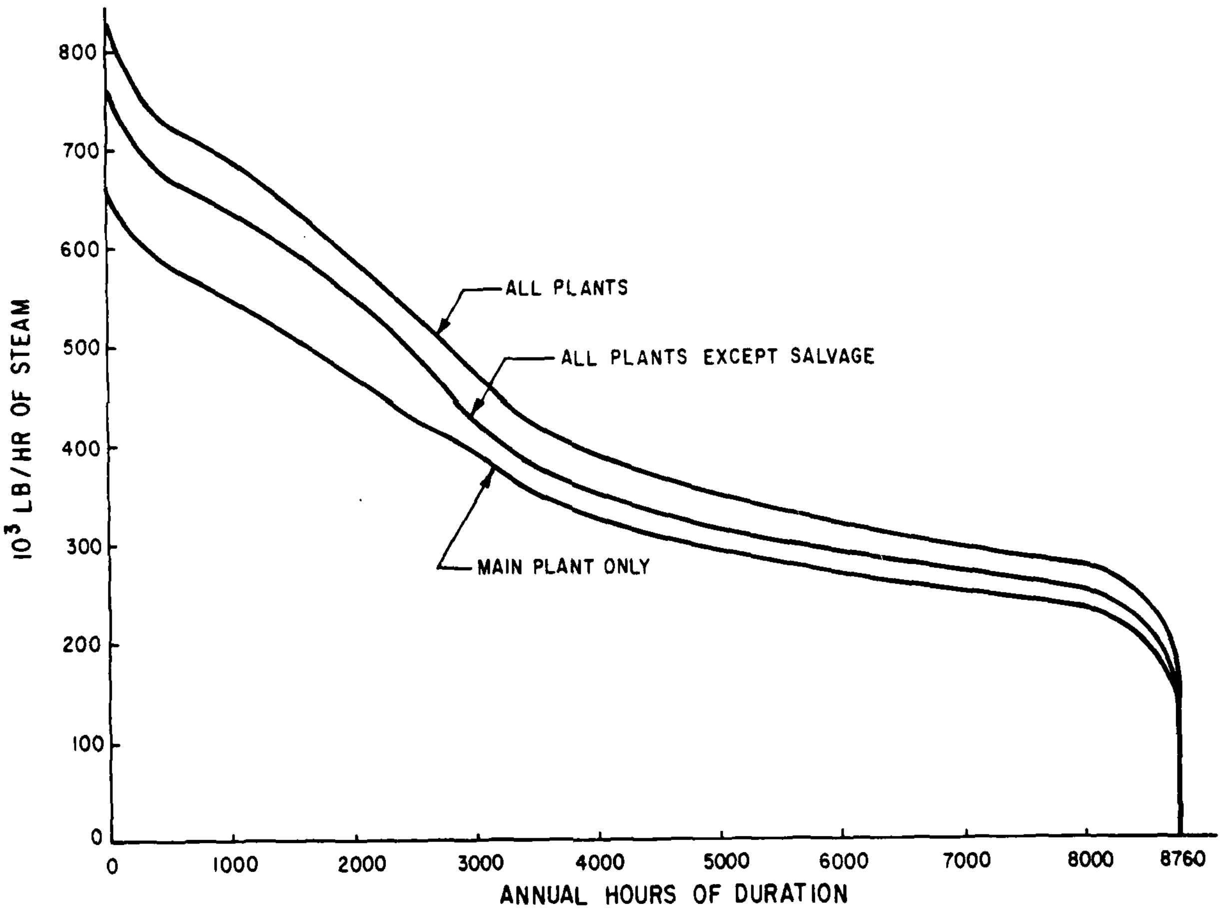


EXHIBIT 1-2

- Main Plant 80-85% of steam load.
- Salvage Plant 4-12% of steam load.
- Peaking Plant 8-14% of steam load.

The steam load distribution curve is especially useful in cogeneration studies. A base steam load which occurs essentially for the full year offers a first indication of magnitude on the size of a feasible cogeneration system. It is seen that a steam load in the range of 270,000 to 290,000 lb/hr occurs for approximately the full year.

The electric load duration curve, Exhibit 1-3, establishes annual electric consumption at approximately 600 x 10^6 MWh per year. When an electric load duration curve, as in the exhibit, indicates there is a certain electric demand occuring throughout the year, another suggestion for magnitude of the size of a cogeneration system is suggested. Here we see that an electric load in the range of 50-60 MW occurs for approximately the full year.

Exhibits 1-4, 1-5 and 1-6 present typical hourly profiles of daily steam and electric loads for a winter day, a spring and fall day and a summer day. It is examination of such hourly profiles which furnishes the most valuable insight in a study such as this. It is seen that while the magnitude of the steam load varies with the season, the steam load is essentially constant over any day. The electric load, however, shows substantial peaks during daytime. These peaks are more pronounced during summer days because of air conditioning requirements. These daily profiles support the conclusion that over the entire year there are coincident loads in the order of 50-60 MW of electric demand and 270,000 lb/hr of steam demand. These profiles will be used to investigate load following requirements for the various cogeneration systems to be studied.

ELECTRIC LOAD DURATION CURVE

PROJECTED TO 1988

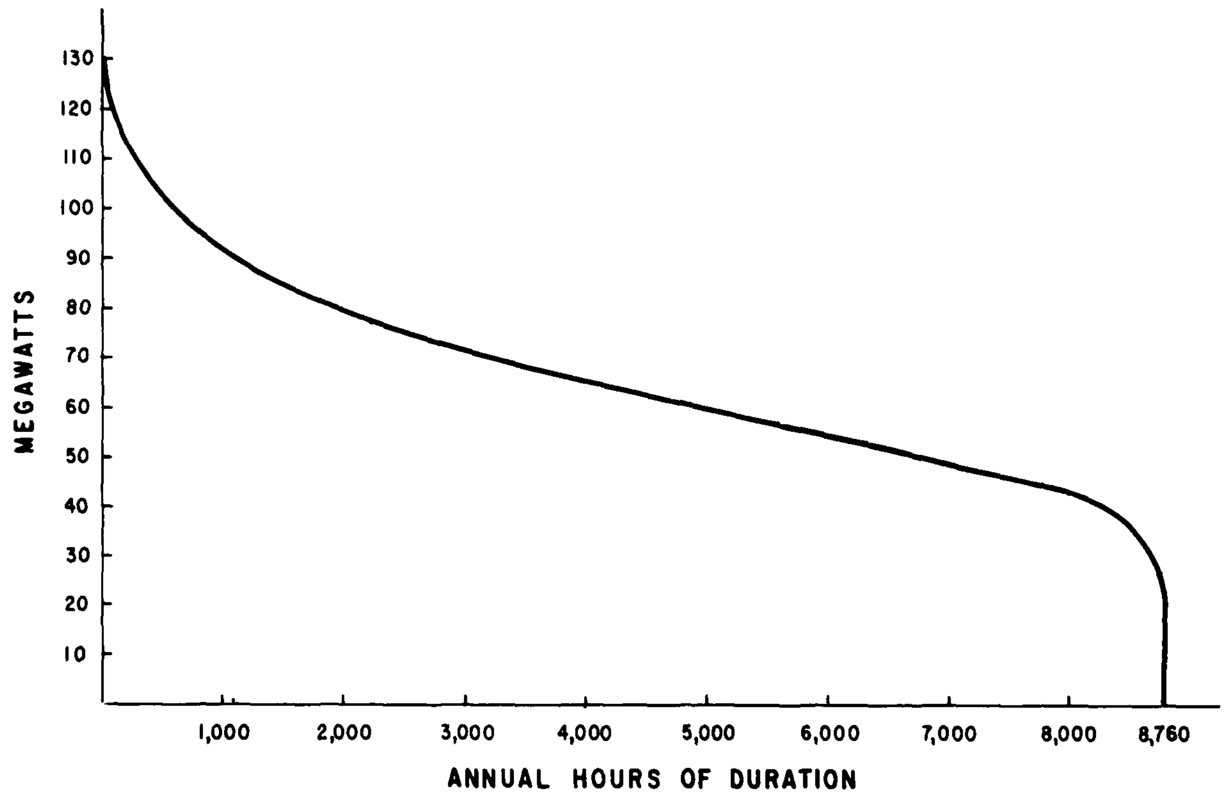


EXHIBIT 1-3

HOURLY PROFILE TYPICAL WINTER DAY STEAM AND ELECTRICAL DEMAND

PROJECTED TO 1988

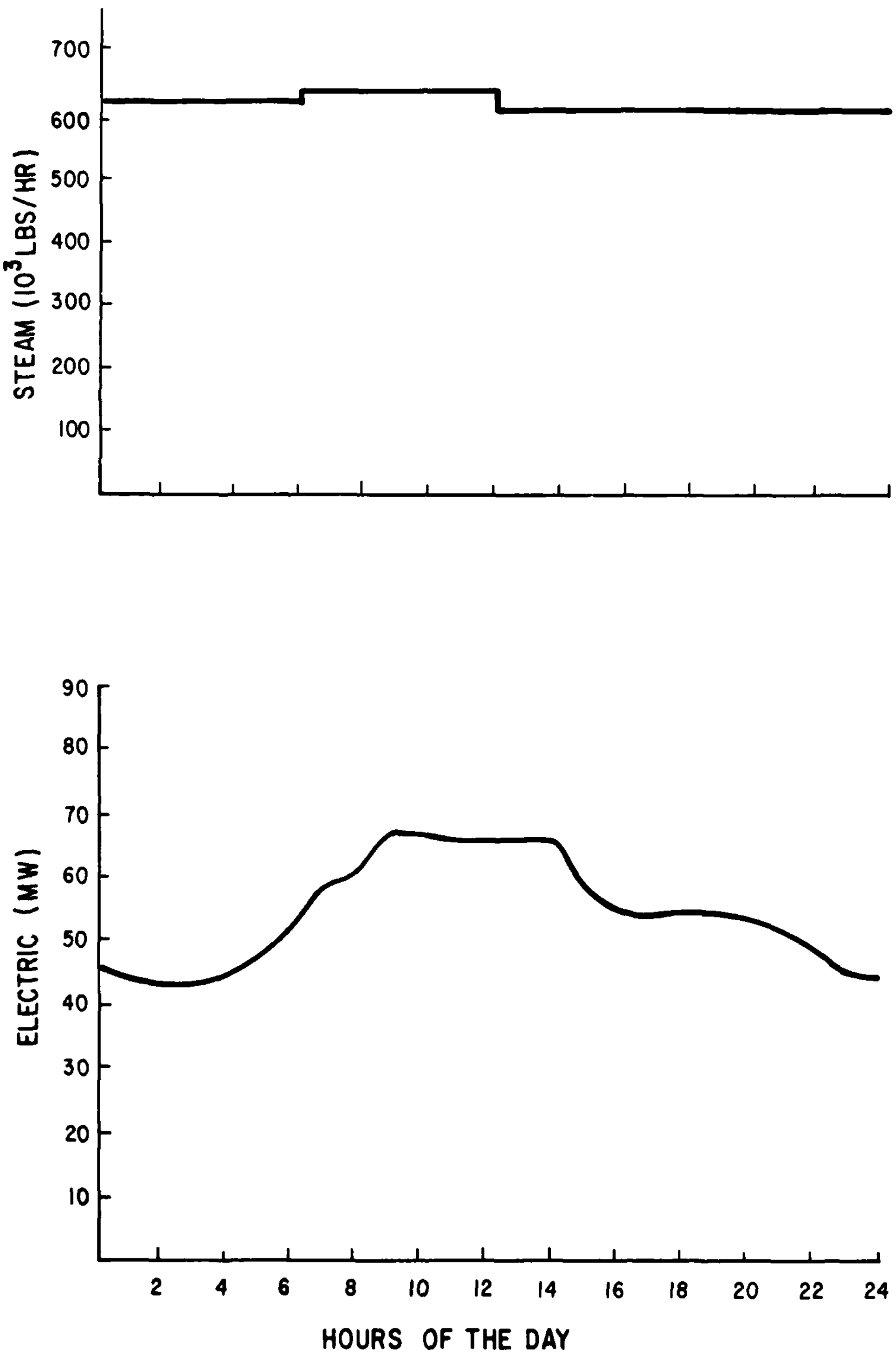


EXHIBIT I-4

HOURLY PROFILE-TYPICAL FALL AND SPRING DAY STEAM AND ELECTRICAL DEMAND

PROJECTED TO 1988

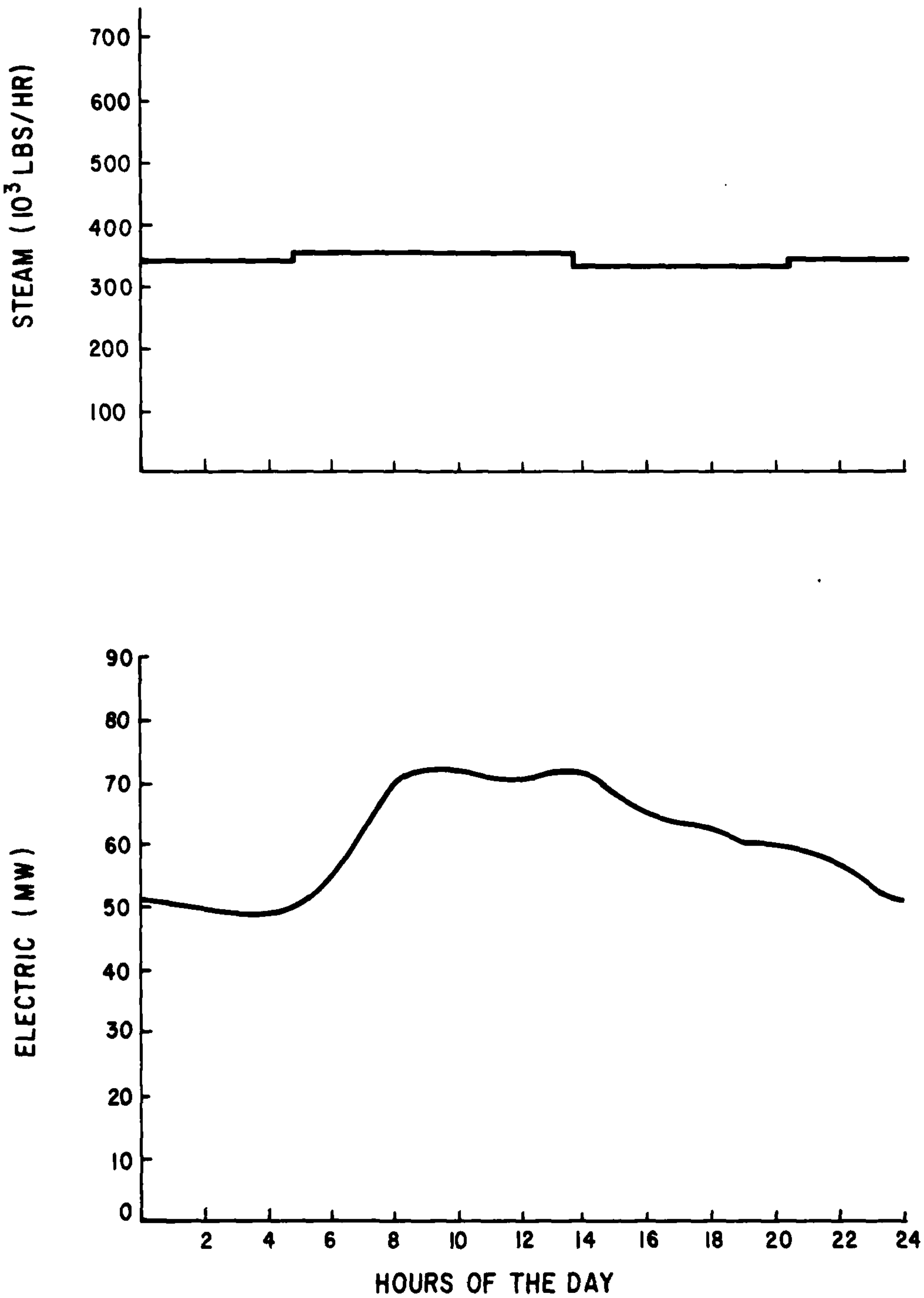


EXHIBIT I-5

HOURLY PROFILE-TYPICAL SUMMER DAY
STEAM AND ELECTRICAL DEMAND

PROJECTED TO 1988

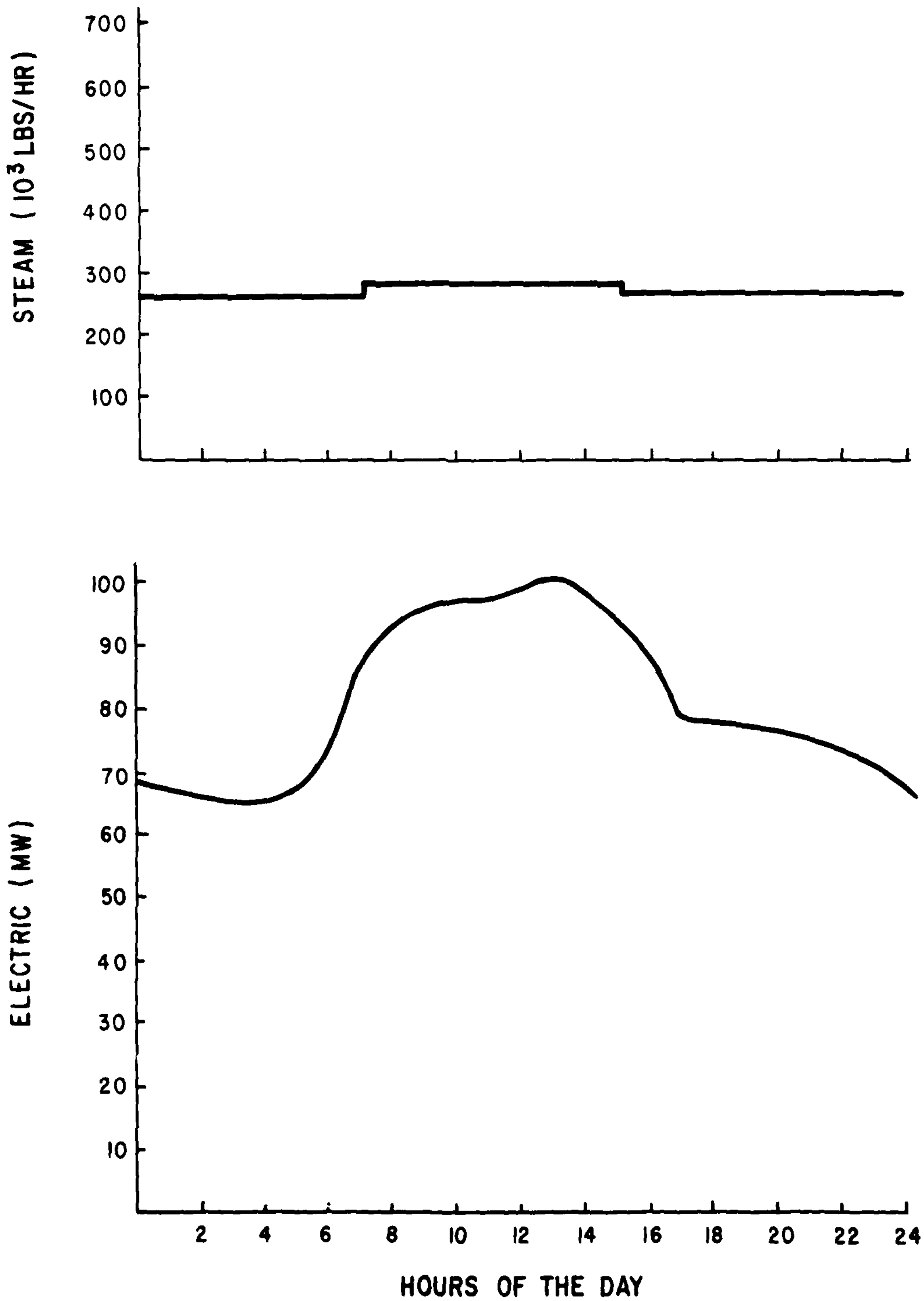


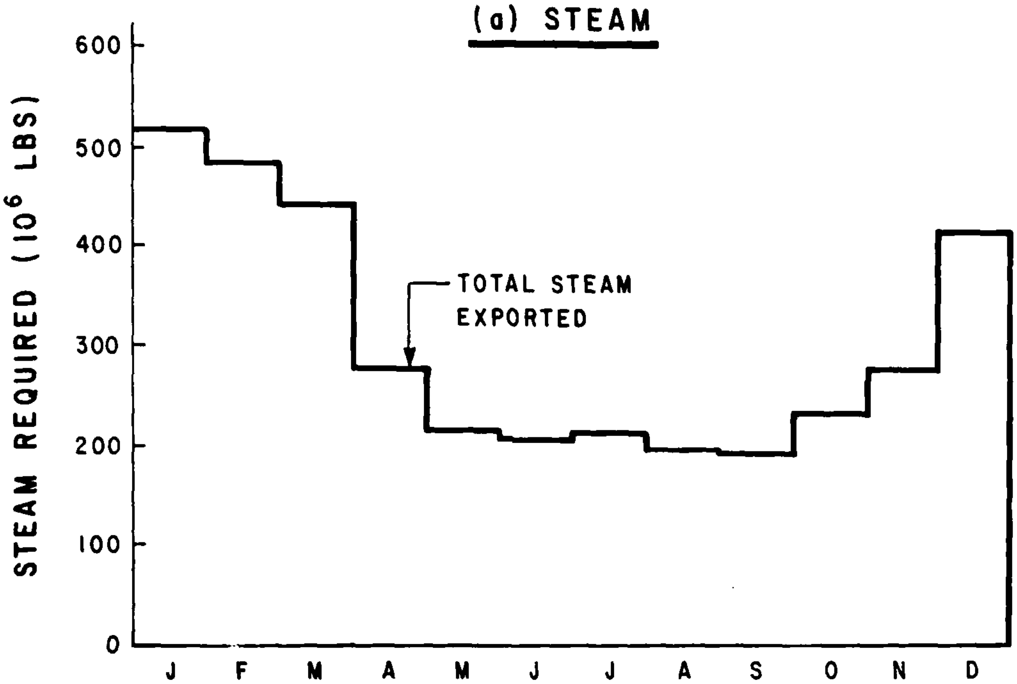
EXHIBIT 1-6

Exhibit 1-7 presents, for general information, the monthly steam and electric consumptions. These reveal the "mismatch" to be expected between overall steam and electric demand: months with high steam requirements have low electric requirements and vice versa. Maximum and minimum monthly steam consumptions are 520 x 10^6 lb/hr and 190 x 10^6 lb/hr respectively, while those for electric are 62 x 10^6 kWh and 36 x 10^6 kWh.

Exhibit 1-8 sets forth averaged monthly electric and steam loads as well as electric peaks. In the usual case, these types of data are not generally useful because of the possible non-coincidence of the steam and electric loads. However, because of the essentially constant nature of the steam load over a twenty-four hour period, this data will prove useful for detailed cogeneration system selection and analysis. This is especially so since there exists a cogeneration rate schedule from the public utility (see Section 6.0) allowing for the power purchase from and sale to the utility. The power to be purchased and/or sold, the capacity and distribution demand charges can be quickly determined for any number of possible sizes and steam/electric mixes of cogeneration system candidates.

Exhibit 1-9 presents steam and electric loads and their ratios in energy terms. These ratios are also essential when cogenerating systems are studied: matching of system outputs to requirements is crucial for economic assessment (see References 10 and 11 for a complete discussion of this point). To take full benefit from these systems, therefore, the ratios featured in this exhibit can be used to choose and compare cogenerating systems designed to supply the entire SPNC loads as well as to select their operating conditions to follow the loads.

PROJECTED LOADS FOR 1988



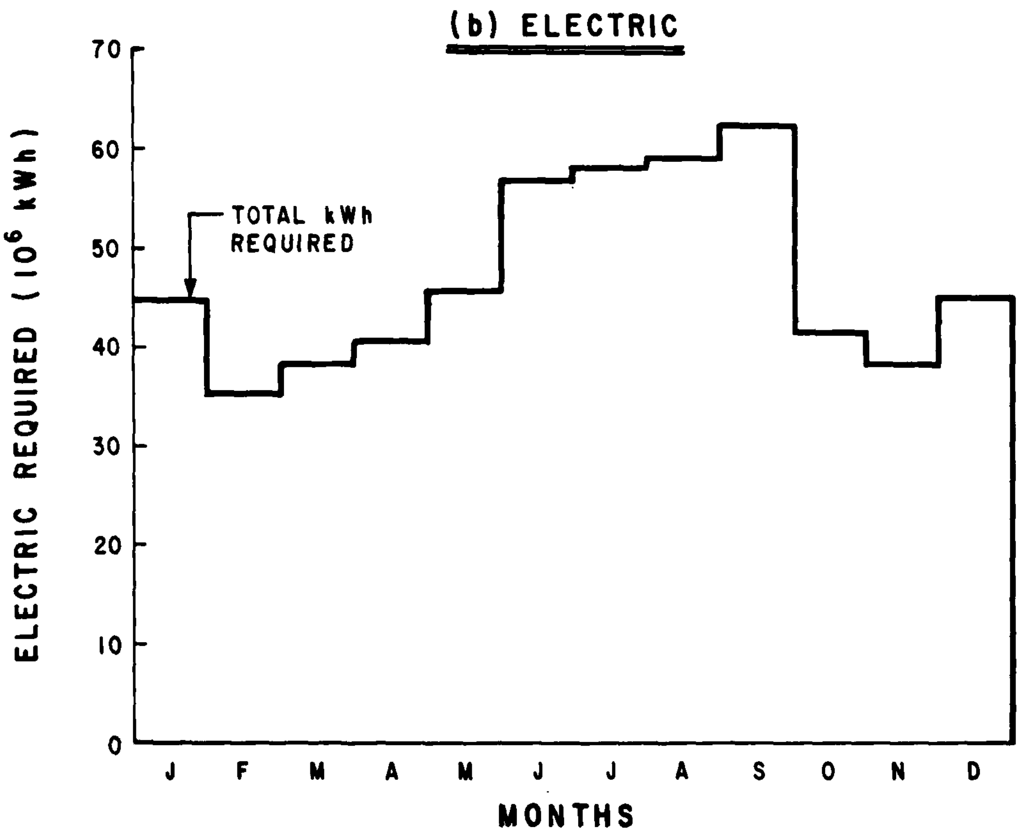


EXHIBIT 1-7

PROJECTED MONTHLY LOADS FOR 1988

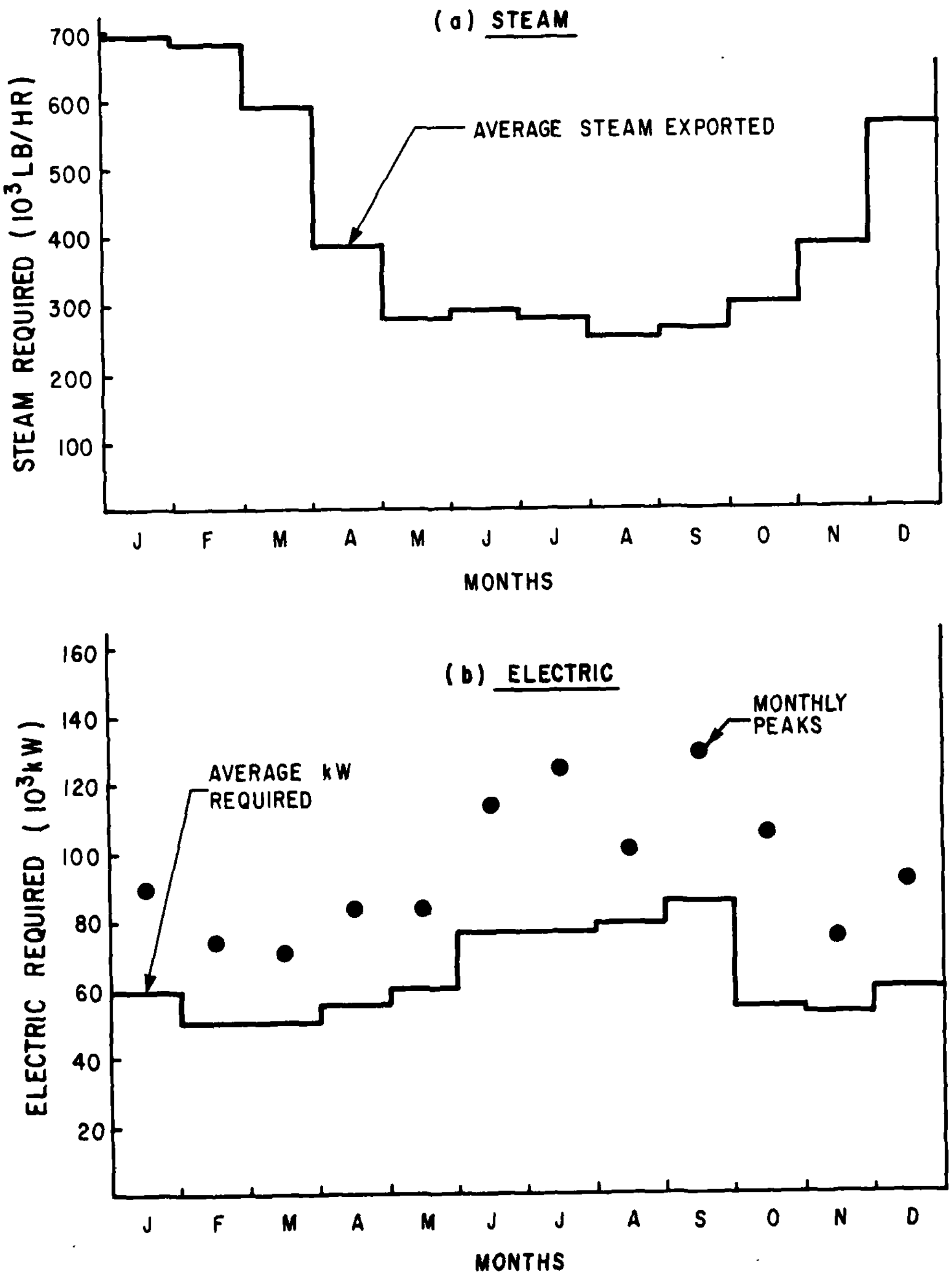


EXHIBIT 1-8

STEAM AND ELECTRIC ENERGY PROFILES

PROJECTED TO 1988

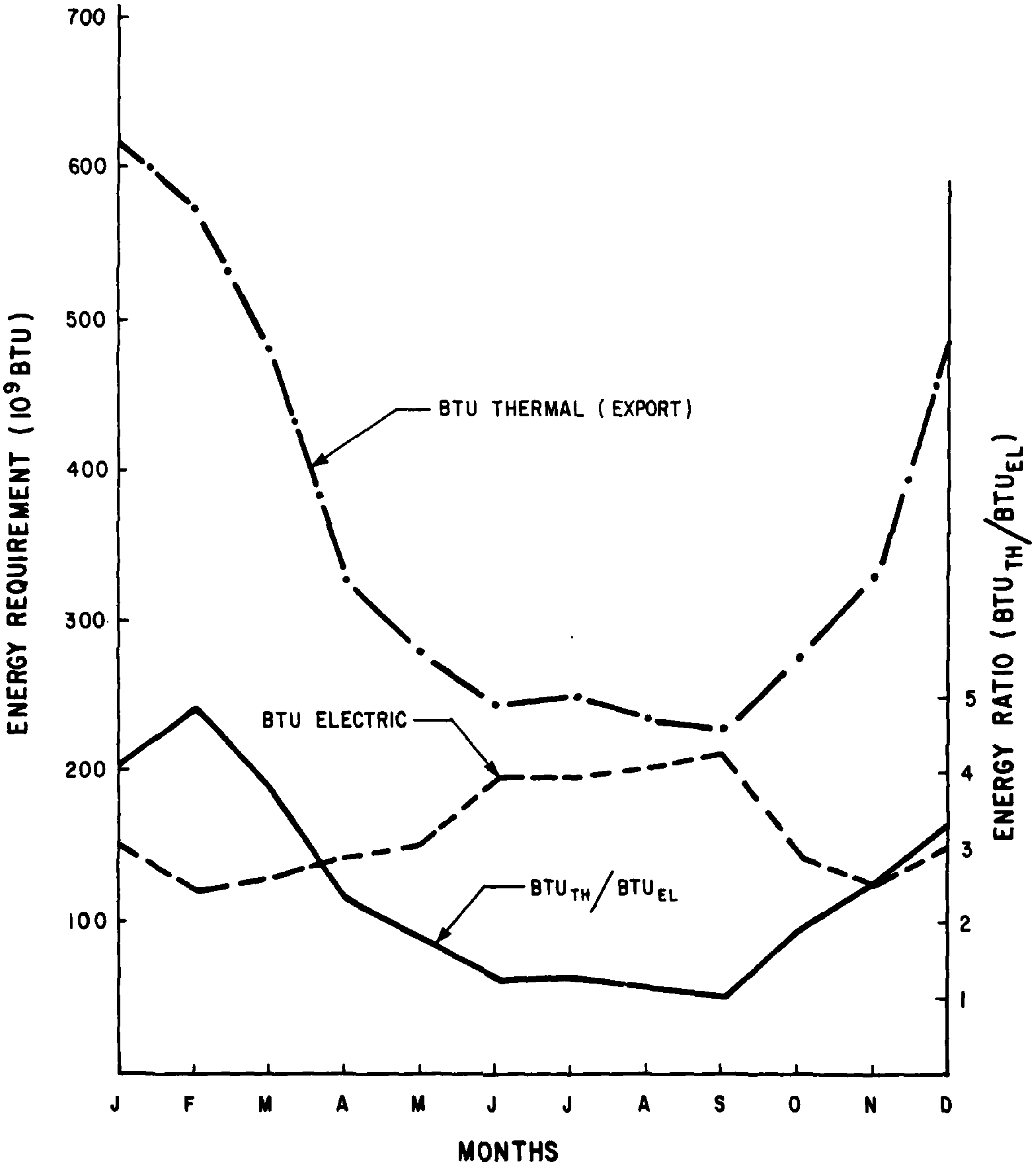


EXHIBIT 1-9

1.3 Coal Availability

To complete this discussion of the facility, an analysis of coal availability has been made. The Defense Fuel Supply Center was contacted to provide current data for Sewells Point. Their response is shown in Table 1-3 which provides properties and costs for both high and low sulfur coals. While the properties show some sizable variations, most of the gasifiers can use an array of coals without regard to particular values or strict specification. Further note that the transportation costs for the high sulfur coals are not shown; they may be estimated from their low sulfur counterparts and from the current literature.

Based on this data and for the purposes of this study, we establish here the following generic coals with associated properties:

	Low Sulfur	High Sulfur
Cost ($/ton delivered)	$ 56	$ 51

Properties

Btu per lb (dry)		13,500 - 14,500
Fixed Carbon	(%)	45 - 50
Volatile Matter	(%)	35 - 40
Ash	(%)	8 - 15
Moisture	(%)	5 - 7
Sulfur	(%)	3.0 - 4.5
AST		2500°F
Hargrove Index		40 - 80
Size		As Mined

TABLE 1-3

SEWELLS POINT NAVAL COMPLEX
COAL AVAILABILITY

(SOURCE: Defense Fuel Supply Center, October 1980)

Coal	Moisture (%)	Volatile Matter (%)	Ash (%)	Sulfur Content (%)	Heating Value (Btu)	Ash Softening Temperature	Hardgrove Grindability Index	Freight Costs ($/ton)	Mine Price ($/ton)
Norfolk, Va. Coal Mountain #9/W. Va.	4.1	34.1	6.1	0.6	14,510	2,910	56	14.82	43.50
Norfolk, Va. Barle #1, Middlesboro, Va.	4.3	35.6	6.9	0.8	14,043	2,410	54	19.97	37.00
Norfolk, Va. Ambrose Branch, Va.	3.8	33.1	4.8	0.7	14,689	2,720	55	14.82	38.25
Average Low Sulfur Coal	4.1	34.3	5.9	0.7	14,414	2,680	55	16.52	39.58
Rock Island, Ill. Globe Fuel Inc., Marion, Ill.	9.5	36.6	14.8	3.7	12,295	2,100	56	N/A	25.00
Ft. McCoy, Wisconsin Spancer Coal Company, Chrising, In.	10.8	45.3	9.9	5.9	13,010	2,050	N/A	N/A	41.75
Ft. Benjamin Harrison, In. PV Corp., Ireland, In.	13.7	40.9	7.8	3.3	13,508	2,060	53	N/A	31.50
Average High Sulfur Coal	11.3	40.9	10.8	4.3	12,938	2,070	56	N/A	32.75

1.4 REFERENCES

1.1 "Coal Conversion Study" Power Plant, Naval Shipyard, Portsmouth, New Hampshire, Department of the Navy, Northern Division, Naval Facilities Engineering Command; prepared by Pope, Evans and Robbins, Inc., New York, March 1979.

1.2 "Solid Fuel Conversion Study" Naval Air Station, Brunswick, Maine, Department of the Navy, Northern Division, Naval Facilities Engineering Command; prepared by Pope, Evans and Robbins, Inc., New York, November 1979.

1.3 "Operation Optimization" Goddard Power Plant, Naval Ordnance Station, Indian Head, Maryland, Department of the Navy, Chesapeake Division, Naval Facilities Engineering Command; prepared by Pope, Evans and Robbins, Inc., New York, January 1980.

1.4 "Coal Energy Conversion Options for Navy Bases" Civil Engineering Laboratory, Naval Construction Battalion Center, Port Hueneme, California, December 1978.

1.5 "Coal Fired Boilers at Navy Bases" Civil Engineering Laboratory, Naval Construction Battalion Center, Port Hueneme, California, May 1979.

1.6 "Application Potential of Energy Systems at Navy Sites" Civil Engineering Laboratory, Naval Construction Battalion Center, Port Hueneme, California; prepared by Acruex Corporation, Mountain View, California, October 1979.

1.7 "Co-Generation: A Systematic Analyses of Combined Steam and Power Generation", prepared for Portsmouth Naval Shipyard, Portsmouth, New Hampshire, by Pope, Evans and Robbins, New York, New York, May 1977; see also AIAA Journal of Energy, February-March 1978.

1.8 "Purchased and Generated Electric Power", U.S. Naval Base, Philadelphia, Pennsylvania, prepared by Pope, Evans and Robbins, New York, New York, March 1979.

1.9 "The Development of the Basis for an Intergrated Alternative Energy Plan for the Sewells Point Naval Complex, Norfolk, Virginia" Department of the Navy, Atlantic Division, Naval Facilities Engineering Command, prepared by Batelle's Colombus Laboratories, Columbus, Ohio, November 1977.

1.10 "Study and Evaluation of Energy Requirements and Optional Approaches Including Cogeneration for Supplying Required Energy at the Naval Weapons Station, Yorktown, Virginia" Department of the Navy, Atlantic Division, Naval Facilities Engineering Command; prepared by Sanderson and Porter, Inc., New York, April 1980.

1.11 Gorges, H., "A Cogeneration Plant for the Naval Air Station, Memphis, Tennessee" paper presented at the Integrated Energy Systems Task Group, Washington, D.C., June 9, 1981.

2.0 GASIFICATION PROCESSES

The major thrust of this study is an investigation of coal gasification. Further, this is the first Navy site-specific application of this emerging technology. Therefore, it seems worthwhile to devote a considerable effort to exposing the various elements of the coal gasification process.

Coal gasification is not new, but it is an evolving technology. Atmospheric pressure gasifiers were constructed and used in Europe about 1840. Few industrial applications were made until the 1860's, but by 1880, equipment for cooling and cleaning the hot raw gas was developed in England so that it became possible to use the gas in small furnaces and gas engines. There were about 150 companies in Europe and the United States building gasification plants in the early 1900's. At that time there were about 500 gas engine installations in the United States. In addition to engines, the gas was used for heating furnaces and kilns in the steel and glass industries, in ceramics and lime-burning plants, as well as in other metallurgical and chemical fields.

In 1921 there were about 11,000 gasifiers in use in the United States. These gasifiers consumed more than 15 million tons of coal a year. In the early 1920's competition from petroleum and natural gas products resulted in a rapid decline in the number of gasifiers in use; however, in 1948 there were still about 2,000 gasifiers in use. Since 1948, the number has diminished so that no significant number of gasifiers are now in use.

However, with the continuing shortage of petroleum and natural gas and with their escalating costs, renewed interest in gasification processes has been generated since the early 1970's. In recent years emphasis has been on establishing and developing commercially available equipment.

Coal gasification is the broad term used to describe the conversion of coal to gas. Within the broad spectrum there is a classification of the product gas in terms of its end use. This potential market has generally been divided into three areas:

(i) High-Btu gas - a substitute natural gas with heating value above 900 Btu/scf,

(ii) Medium-Btu gas - a fuel gas with a heating value between 200 and 400 Btu/scf, requiring oxygen as the oxidant,

(iii) Low-Btu gas - a fuel gas with a heating value below 200 Btu/scf, requiring air as the oxidant.

The focus of this study is on low and medium Btu gas processes. A generic flow diagram exposing the basic process steps is shown in Exhibit 2-1.

The remainder of this section first provides a general description of standard gasifier types with comparison between them. Details of representative and commercially available systems follow. Finally, for the several manufacturers, comparison of pertinent performance quantities are provided.

2.1 General Processes

Coal gasification processes are categorized according to the techniques in which the various reactants -- coal, steam, oxidant (air or oxygen) -- contact each other and according to the movement of the coal itself. In general, we address three types: fixed bed, fluidized bed and entrained bed. Simplified comparisons of these configurations are shown in Exhibit 2-2, which should be referred to during the following discussion, derived from References 1, 2 and 3.

GENERIC COAL GASIFICATION FLOW DIAGRAM

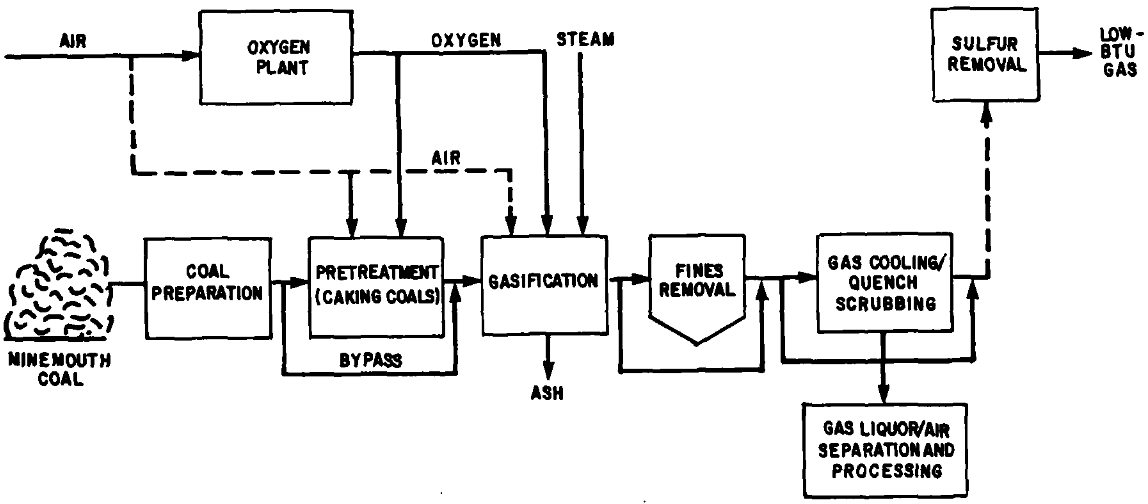


EXHIBIT 2-1

COAL AND OXIDANT FLOW IN COAL GASIFIERS

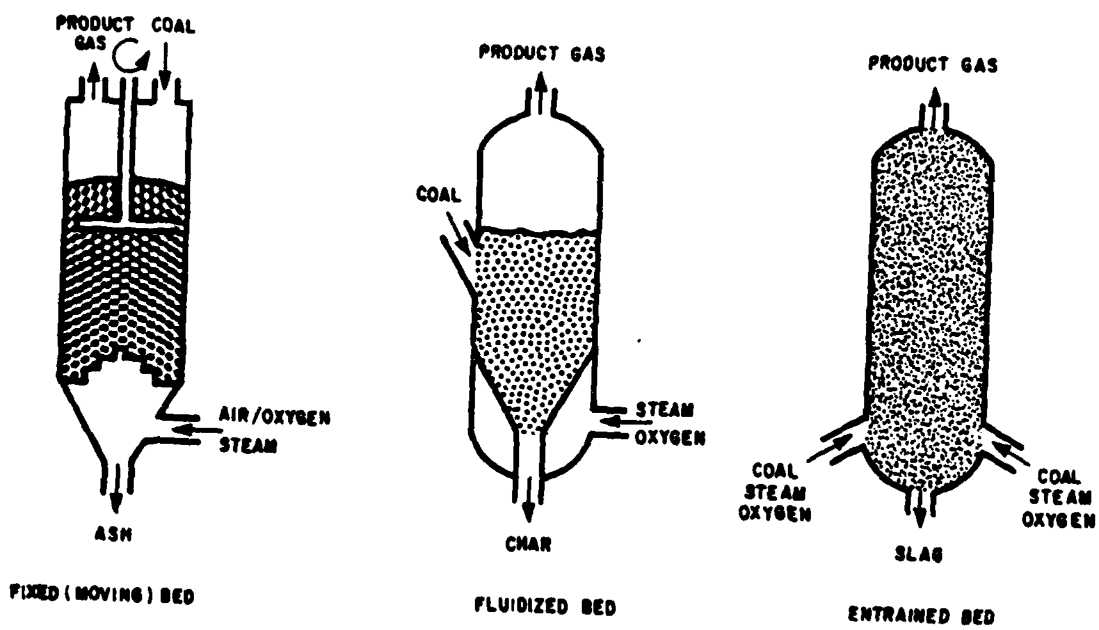


EXHIBIT 2-2

Consider first the fixed bed process. Because the flow of coal and residue (ash) is countercurrent to the gasifying agents and products (principally carbon monoxide and hydrogen), fixed beds exhibit excellent thermal efficiencies. For example, the outgoing ash heats the incoming gases, and the outgoing products heat the incoming coal. Moreover, the long residence times of coal particles moving through the bed allow high carbon conversion efficiencies.

Within a fixed bed are various zones of progressively higher temperatures to which the incoming coal is subjected. These zones are:

- Drying Zone: Raw coal (sized 1/4 to 1-1/2 inch) fed to the reactor comes in contact with the hot product gases, and moisture in the coal is driven off.

- Devolatilization Zone: As the coal is heated further, occluded carbon dioxide and methane are driven off at temperatures less than 400°F. Organic sulfur in the coal is decomposed in the range of 400°F to 900°F and is converted to hydrogen sulfide and other compounds. Nitrogen compounds in the coal decompose to release nitrogen and ammonia. Above 550°F, oils and tars are distilled from the coal.

- Gasification Zone: Char (the now-devolatilized coal) comes in contact with steam and the hot combustion products from the zone directly below. The chief reactants here are that of carbon monoxide and hydrogen being formed from the combination of carbon with water and carbon dioxide. These reactions are endothermic, and the production of carbon monoxide and hydrogen are favored at high

temperatures; whereas, the production of carbon dioxide and hydrogen would be favored at lower temperatures.

- Combustion Zone: This zone, which supplies both the heat and carbon dioxide for the gasification zone, consists of a layer of ash physically supporting the combusting (now gasified) char. The key reaction in this zone is that of carbon with oxygen, which produces heat and carbon dioxide.

- The Ash Bed: Located at the bottom, the ash bed acts as a distributor for the oxygen (or air) and steam and, more importantly, provides heat to incoming feeds.

Fixed-bed gasifiers can be further classified into single- and two-stage units. Both types will contain the zones described above; they differ in the location of gas removal and the temperature ranges within the devolatilization and drying zones. A single-stage gasifier has only one product gas offtake, at the top of the coal bed above the drying zone. Typical temperatures of gas leaving the unit are in the range of 700° to 1100°F. Thus, incoming coal is heated very rapidly and causes the oils and tars from the coal to crack and polymerize to heavy viscous tar and pitch. This violent distillation also causes the coal to decrepitate and gives rise to coal dust, which is carried out with the product gas.

Two-stage producers have one gas offtake above the drying zone and one just at the top of the gasification zone, where about half the gas produced by gasification is removed; the remainder flows upward through the devolatilization and drying zones. The temperatures attained in these two zones

are considerably lower than those seen in single-stage units. Therefore, the incoming coal is heated, and the oils and tars are evolved in a much slower manner; thus, the problems in handling heavy tars, pitch and soot are avoided.

Next, consider the fluidized bed process. A stationary bed of coal becomes fluidized when the pressure drop of the gasifying agents moving through the bed is sufficient to lift the coal particles. This requires smaller coal sizes than the fixed-bed units, normally in the 10 to 100 mesh (0.078-0.0058 inch) size range. At this stage, the bed expands, and the coal particles move about randomly. This fluidized action causes thorough mixing of the coal and the gases, and the bed exhibits almost isothermal conditions (variations are typically < 100°F). Bed temperatures characteristically are in the 1500°-1900°F range, depending on coal type.

Because of these mixing properties, fluidized beds can handle a higher coal feed rate than can fixed beds for the same size reactor. The temperature of the reactor exit gases is about the same as that of the bed, and a heat exchange device is required to economize heat. Compared to fixed beds, fluidized beds have, in general:

- more solids carried over with the product gases,
- less tar and soot production, and
- more unreacted carbon in the ash.

Ideally, ash removal would be accompanied by the heavier ash particles working their way through the bed and falling out at the bottom. In the only commercially available fluidized-bed gasifier, the Winkler gasifier, about 30% of the ash is removed in this manner; the remaining 70% is carried out with the product gases (see Reference 4).

Finally, we provide a general description of entrained bed processes (see References 5 and 6). As opposed to a fixed bed in which coal particles move counter-currently to the reactive and product gases, and as opposed to a fluidized bed in which most of the coal particles are suspended by the gaseous phase, the particles in an entrained gasifier are carried, or entrained, by the reacting gases. The coal particles used in entrained gasifiers, therefore, are usually much smaller than those used in other systems because large particles would require excessive gas velocities, and because higher conversion rates are needed because of the shorter solids residence time. The coal is usually pulverized to a normal plant grind of 70 percent through 200 mesh.

The reactants -- coal, oxygen/air, steam -- are typically introduced into the gasifier at high velocity through one or more burners, or nozzles. The high velocity is required to prevent the flame front from retreating into the nozzle itself -- a condition known as flashback. The burners are usually composed of concentric pipes with one or more of the reactants flowing through each pipe and mixing at the burner tip. The burners can be oriented in the gasifier in many ways, including tangentially, radially opposed, and axially. The performance of a gasifier can be affected to a large degree by the flow characteristics and mixing efficiency of the burners. Great care is normally used to align the burners in a way which minimizes the impingement of the high velocity reactants on the gasifier surfaces.

Flame temperatures at the burner discharges can be as high as 3500°F. This results in the extremely rapid conversion of the coal particles and the destruction of virtually all the higher hydrocarbon species. Outside the immediate flame

regions, heat losses, further mixing with steam, and endothermic reactions combine to lower the gas phase temperature to less than 3000°F. Because of these high reaction temperatures, the oxygen consumption is usually higher and the steam consumption is usally lower than for other gasification systems. The high reaction temperatures also result in the melting of a significant portion of the coal ash so that it is removed from the reaction zone as a liquid slag. The principal species leaving the reaction zone are CO, CO_2, H_2, N_2 (if air is used), and unreacted steam and char. Most of the sulfur in the coal appears in the gas as H_2S, and as smaller amounts of COS. Usually, at least 70 percent carbon conversion can be easily achieved with a single pass.

The typical range is probably 80 to 95 percent conversion at gas and solid residence times of several seconds. Nearly 100 percent conversion can be achieved if the char is recycled to extinction, since the only losses would occur as carbon trapped in the slag and as carbon lost or not captured by the recycle equipment. Due to the processing conditions, almost all coals can be used in an entrained gasifier without the need for oxygen pretreatment or the concern for agglomeration associated with other gasification schemes.

A survey of the basic characteristics of the several types of gasifiers are shown in Tables 2-1, 2-2 and 2-3 for fixed, fluidized and entrained beds respectively (see also Reference 3). Combining these results lead to the display in Table 2-4 where basic advantages and disadvantages are shown. Detailed comparisons of commercially available gasifiers in each category are provided later.

TABLE 2-1

FIXED BED GASIFIER CHARACTERISTICS

Characteristics	Advantage	Limitation
Experience	A mature tecnology with many commercial designs available, including pressurized conditons.	
Complexity		Internal moving parts with higher degree of mechanical complexity.
Capacity		Lowest capacity for three generic processes due to limited gas-flow rates.
Inventory (Coal)	Large fuel inventory provides safety, reliability, stability; fixed bed may be banked for long periods.	
Feed Coal Type		Commercial operation with caking coal is less certain and requires agitation at lower throughout.
Handling		Sized coal is required and fines must be disposed of or handled separately.
Product Gas		Contains tar and oil, phenols, ammonia, and a small amount of dust; more complicated multistage cleanup required; low temperature precludes heat recovery.
Ash Removal	Dry ash contains little carbon due to long residence time and countercurrent flow of solids and gases.	
Temperature	Gradient in temperature (due to countercurrent flow) decreases thermal losses; lower temperature operation avoids problems with material of construction.	Control of temperature below ash fusion point is required to avoid clinkering; steam injected for this purpose leads to thermal losses and cleanup problems.
Operating Range	Highest capability for turndown.	

TABLE 2-2

FLUIDIZED BED GASIFIER CHARACTERISTICS

Characteristics	Advantage	Limitation
Experience	Commercial design available and extensive development under way on second-generation processes.	Less developed than fixed bed and poorer market acceptance for commercial model.
Complexity	Less complex than fixed bed and has no internal moving parts.	Design of distributor plate is important for successful operation.
Capacity	Higher capacity due to lower residence time.	Capacity limited by entrainment at high gas velocities.
Inventory (Coal)	Large fuel inventory provides safety, reliability, and stability.	Inventory of coal is reduced by the high content of inert materials in the bed.
Feed Coal		
Type	Handles a wide variety of fuels.	Caking coal requires pretreatment.
Handling	Handles a wide variety of particle sizes.	Removal of fines required to prevent elutriation or flow instability.
Product Gas	Less tars and phenols; unvarying composition due to uniform conditions in bed; excellent solid-gas contact.	Carryover of ash and char is high; sensible heat loss is moderate, but reduction of temperature would lead to tar formation.
Ash Removal	Dry ash gravitates to bottom of bed for removal.	More carbon leaves in ash due to uniform composition of fluid bed.
Temperature	Gasifier can operate at lower temperature to reduce thermal losses; steam injection or water jacket is not generally required for cooling.	
Operating Range	Moderate turndown capability.	Range is limited by gas velocity required to maintain fluidization at acceptable rate of elutriation.

TABLE 2-3

ENTRAINED FLOW GASIFIER CHARACTERISTICS

Characteristics	Advantage	Limitation
Experience	Commercial designs available and development program exists on second-generation processes.	Less developed than fixed bed.
Complexity	No moving parts and has simpler geometry than fluid bed. Water jackets add to system complexity.	Critical design areas include combustor nozzles and heat recovery in presence of molten slag.
Capacity	Highest capacity per unit volume.	
Inventory (Coal)		Smallest inventory of three generic classes; requires advanced control techniques to ensure safe, reliable operation.
Feed Coal Type	Any coal may be used without pretreatment.	
Handling	No fines are rejected.	Pulverizing and drying of surface moisture are required. Potential erosion due to gas-solid streams.
Product Gas	Free of tars and phenols.	Ash, char and sensible heat in gas must be recovered, which reduces efficiency.
Ash Removal	Produces inert slagged ash with low content; fines carried over can be recycled to gasifier.	Higher thermal loss in ash.
Temperature		Highest temperature of three classes (1) causes thermal losses, (2) requires better materials of construction, and (3) requires greater use of oxygen or preheated air which results in higher CO_2 content in product gas.
Operating Range		Process has the least operating range and is limited by need to maintain slagging conditions without degrading refractories.

TABLE 2-4

COMPARISON OF GENERIC GASIFICATION PROCESSES

A. ADVANTAGES

FIXED BED	FLUIDIZED BED	ENTRAINED BED
• High Carbon Conversion Efficiency • Low Ash Carryover • Low Temperature Operation • Lowest Air/Oxygen Requirement	• High Degree of Process Uniformity • Excellent Solid/Gas Contact • Lower Residence Time Than Fixed Bed Gasifier • Higher Coal Throughout Per Unit Volume of Reactor	• Handles all Types of Coal - No Pretreatment • Low Steam Consumption • Excellent Solid/Gas Contact • No Tar Formation • No Phenol Formation • Ability to Slag Ash • High Capacity Per Unit Volume of Reactor • Produces Inert Slagged Ash

B. DISADVANTAGES

FIXED BED	FLUIDIZED BED	ENTRAINED BED
• Sized Coal Required • Coal Fines Must Be Briquetted • Low Capacity • Low Offgas Temperature • Produces Tars and Heavier Hydrocarbons • High Steam Consumption • Produces Phenols • Use of Caking Coals Not Commercially Proven	• Sized Coal Required • Dry Coal Required For Feeding • Requires Complicated Gas Distributor • Caking Coals Require Pretreatment • High Carbon Loss With Ash • Fluidization Requirement Sensitive to Fuel Characteristics	• Requires Finely Crushed Coal 7.0% < 200 Mesh • Small Surge Capacity Requiring Close Control

2.2 Commercially Available Gasifiers

Various gasifiers falling into the three generic categories were screened according to several factors, to be provided below, so that a small representative set might be established for detailed performance analysis and economic evaluation. An overall list of gasifiers is shown in Table 2-5.

The major categories of screening are described below:

Status - This factor pertains to the degree of development or commercialization. Those processes that were commercial or were thought to become commercially available by the time of facility design were favored.

Technology Factors - These included complexity, feed coal types, operating experience and conditions and conversion efficiency. Considerations here were to favor those gasifiers with moderate or lower complexity, capability to accept a wide range of coal, good operating experiences including maintenance records, and high efficiency. We also sought to include representative processes from the three generic classes.

Capacity - Here we sought to evaluate the number of gasifiers needed to handle selected amounts of coal. Since this is not a utiltity-type operation but rather an industrial gasification application, low to moderate capacity was favored.

Data Availability - Nothwithstanding any of the above factors, data availability in the open literature was considered of prime importance. If the system under evaluation did not have a data base sufficient for cycle assessment, it was deemed unsuitable for this feasibility study.

TABLE 2-5

TOTAL POPULATION OF LOW/MEDIUM-BTU GASIFIERS

GASIFIER TYPE

Gasifier Name	Licensor/Developer	Status
USBM Annular Retort	Bureau of Mines/ERDA (USA)	Past development unit testing; lignite only.
USBM Electrically Heated	Bureau of Mines/ERDA (USA)	Past development unit testing.
Entrained-Bed, Slagging Ash		
Koppers-Totzek	Koppers Company (USA)	Present commercial operation.
Bi-Gas	Bituminous Coal Research, Inc. (USA)	Present development unit testing.
Texaco	Texaco Development Corp. (USA)	Present development unit testing.
Coalex	Inex Resources, Inc. (USA)	Present development unit testing; commercially available.
FAMCO/Foster Wheeler	Pittsburgh and Midway Coal Co./ Foster Wheeler (USA)	Present development unit testing.
Combustion Engineering	Combustion Engineering (USA)	Present development unit testing.
Brigham Young University	Brigham Young University/ Bituminous Coal Research (USA)	Present development unit testing.
Babcock and Wilcox	The Babcock and Wilcox Co. (USA)	Past commercial operation.
Ruhrgas Vortex	Ruhrgas A.G. (West Germany)	Past commercial operation.
IGT Cyclonizer	Institute of Gas Technology (USA)	Past development unit testing.
Inland Steel	Inland Steel Company (USA)	Past development unit testing.
USBM, Morgantown	Morgantown Energy Research Center/ERDA (USA)	Past development unit testing.
Great Northern Railway	Great Northern Railway Co. (USA)	Past development unit testing.

TABLE 2-5 (Continued)

GASIFIER TYPE

Gasifier Name	Licensor/Developer	Status
Fluidized-Bed, Dry Ash		
Winkler	Davy Powergas Company (USA)	Present comercial operation.
Hygas	Institute of Gas Technology (USA)	Present development unit testing.
Synthane	Pittsburgh Energy Research Center/ERDA (USA)	Present development unit testing.
Hydrane	Pittsburgh Energy Research Center/ERDA (USA)	Present development unit testing.
Cogas	Exxon Corporation (USA)	Present development unit testing.
BCR Low-Btu	Bituminous Coal Research (USA)	Present development unit testing.
CO_2 Acceptor	Consolidation Coal Company (USA)	Present development unit testing.
Electrofluidic Gasification	Iowa State University/ERDA (USA)	Present development unit testing.
LR Fluid Bed	Unknown (Germany)	Past commercial operation.
HRI Fluidized Bed	Hydrocarbon Research, Inc. (USA)	Past development unit testing.
BASF-Flesch-Damar	Badioche Anilin und Boda Fabrik (West Germany)	Past development unit testing.
Fluidized-Bed, Agglomerating Ash		
U-Gas	Institute of Gas Technology (USA)	Present development unit testing.
Battelle/Carbide	Battelle Memorial Institute (USA)	Present development unit testing.
Westinghouse	Westinghouse Electric Corp. (USA)	Present development unit testing.
City College of NY Mark 1	Hydrocarbon Research Inc./ A.M. Squires (USA)	Present development unit testing.
Two-Stage Fluidized	British Gas Council (England)	Present development unit testing.
ICI Moving Burden	Imperial Chemical Industries, Ltd. (England)	Past development unit testing.
Entrained-Bed, Dry Ash		
Garrett Flash Pyrolysis	Garrett Research and Development Company (USA)	Present development unit testing.

TABLE 2-5 (Continued)

GASIFIER TYPE

Gasifier Name	Licensor/Developer	Status
Fixed-Bed, Dry Ash		
Lurgi	American Lurgi Corp. (USA)	Present commercial operation.
Wellman-Galusha	McDowell-Wellman Engr. Co. (USA)	Present commercial operation.
Chapman (Wilputte)	Willpute Corp. (USA)	Present commercial operation.
Woodall-Duckham/Gas Integrale	Woodall-Duckham, Ltd. (USA)	Present commercial operation.
Riley Morgan	Riley Stoker Corp. (USA)	Present demonstration unit testing; commercially available.
Pressurized Wellman-Galusha (KEEC)	Morgantown Energy Research Center/ERDA (USA)	Present development unit testing.
Foster Wheeler/Stoic	Foster Wheeler Energy Corp. (USA)	Present commercial operation.
Kilngas	Allis Chalmers Corp. (USA)	Present development unit testing; commercially available.
Kellogg Fixed Bed	M. W. Kellogg Company (USA)	Present development unit testing.
GEGAS	General Electric Research and Development (USA)	Present development unit testing.
Consol Fixed Bed	Consolidation Coal Company (USA)	Present development unit testing.
IFE Two Stage	International Furnace Equipment Company, Ltd.	Past commercial operation.
Kerpely Producer	Bureau of Mines/ERDA (USA)	Past commercial operation.
Pintech Hillebrand	Unknown (Germany)	Past commercial operation.
U.G.I. Blue Water Gas	U.G.I. Corp./DuPont (USA)	Past commercial operation; coke only.
Power Gas	Power Gas Company (USA)	Past commercial operation.
Wellman Incandescent	Applied Technology (USA)	Past commercial operation.
Fixed-Bed, Slagging Ash		
GG/Lurgi Slagging Gasifier	British Gas Council (GB) Lurgi Mineraloltechnik (W. Germany)	Present development unit testing.
CFERC Slagging Gasifier	Grand Forks Energy Research Center/ERDA (USA)	Present development unit testing; lignite only.

On these bases then, the gasifiers shown in Table 2-5 were assessed. From them the following were selected as representative of the commercially available systems:

Fixed Bed	Lurgi, dry ash
	Wellman - Galusha
	Woodall - Duckham
Fluidized Bed	Winkler
Entrained Bed	Koppers-Totzek
	Texaco

We would emphasize that these six may not be the only gasifiers which can fit current state-of-the-art criteria, but rather are representative of the variety of systems that are available. Indeed by the time the design and bidding phases of this project take place, others might also be of interest.

For each of the six gasifiers we provide summary descriptions in this section (Exhibits 2-3 to 2-8) and detailed process discussions in Appendix A. This data has been taken largely from References 7 and 8.

2.3 Process Comparisons

We next compare the six selected gasifiers for a variety of parameters of importance to cycle and system performance for typical coals. Economic assessments are deferred to a later section.

Consider first operating conditions. Comparisons are shown in Table 2-6. Sizing of the coals for the fixed and fluidized bed types is required. The entrained beds, operating on pulverized coal, show an advantage here. Except for the Winkler fluidized bed process, all gasifiers have reasonable input rates per unit, sufficient to allow suitable redundancy at reasonable economic cost. As we will see later, operating at elevated pressure is an advantage for the combined

SUMMARY DESCRIPTION - LURGI GASIFIER

NAME:	Lurgi, Dry Ash
DEVELOPER/ LICENSOR	American Lurgi Corporation 377 Route 17 Hasbrouck Heights, NJ 07604
TYPE:	Pressurized fixed bed.
STATUS/ HISTORY:	Commercial proven since 1936; eighteen commercial plants have been installed worldwide (outside the U.S.).
CONDITIONS:	Pressure: 350 to 450 psig. Temperature: 1800 to 2500°F combustion zone, 1150 to 1500°F gasification zone, 700 to 1100°F exiting gas. The operating temperature is strongly dependent on the coal type. Expected turndown ratio is 100:0 (American Lurgi).
REACTANTS:	Sized coal (1/8 to 1-1/2 inch). Steam: 3.2 lb per lb of coal. Oxygen: 0.6 lb per lb of coal (Pittsburgh No. 8). Oxidant and steam consumption are dependent on the coal type. Air can be used.
PRODUCTS:	Medium Btu Gas: Oxygen blown (60 to 70 Mscf per ton of coal at 285 to 300 Btu/scf). Low Btu Gas: Air blown (100 Mscf per ton of coal at 179 Btu/scf). By-Products: Tar, tar oil, naphtha, gas liquor, steam, phenols, sulfur and ammonia.
FEED METHODS:	Gravity-fed from coal lock hopper.
ASH REMOVAL:	Dry ash collected in an ash lock hopper.
DESCRIPTION:	Coal is fed downward over a mechanical distributor into a vertical cylindrical, water-jacketed shell. Steam and oxygen (or air) are fed upward through a rotating grate on which the falling ash collects. Ash is removed at the bottom. Raw product gas escapes at the top and is sent downstream for treatment. The gasifier can handle caking coals if mechanical stirring is provided.

EXHIBIT 2-3

SUMMARY DESCRIPTION - WELLMAN-GALUSHA GASIFIER

NAME:	Wellman-Galusha
DEVELOPER/ LICENSOR:	McDowell-Wellman Company 113 St. Clair Avenue, NE Cleveland, Ohio 44114
TYPE:	Fixed bed with or without central agitator.
STATUS/ HISTORY:	Twelve gasifiers operating in U.S., several more overseas. DOE project in Morgantown has been operating for over nine years. Commercial operation: 35 years for Wellman-Galusha design and over 80 years total for all McDowell-Wellman designs.
CONDITIONS:	Pressure: Atmospheric. Temperature: Combustion Zone = 2400°F; gas leaving = 1100 to 1200°F (bituminous) or 600 to 1000°F (anthracite).
REACTANTS:	Steam = 0.4 to 0.7 lb per lb of coal; Air = 3.5 lb per lb of coal; Crushed coal +3/16 to 9/16 in. (anthracite) or +1 to 2 inch (bituminous); Agitated gasifier can handle caking bituminous coal.
PRODUCTS:	Low Btu gas (120 to 168 Btu/scf); Medium Btu gas for oxygen-blown operation (270 to 290 Btu/scf).
FEED METHODS:	Gravity fed (controlled by slide valves) from coal bin on top of the gasifier.
ASH REMOVAL:	Withdrawn through eccentric grate to ash cone.
DESCRIPTION:	Expected turndown ratio is 100:25. Capacity of agitated gasifier is about 25% higher than that of gasifier without central agitator. Water-jacketed and brick-lined gasifier models are available.

EXHIBIT 2-4

SUMMARY DESCRIPTION - WOODALL-DUCKHAM GASIFIER

NAME:	Woodall-Duckham
DEVELOPER/ LICENSOR:	Babcock Contractors, Inc. 921 Penn Avenue Pittsburgh, PA 15222
TYPE:	Two-stage, fixed bed process.
STATUS/ HISTORY:	Thirty years producing industrial fuel gases in Milan, Italy; process used about 20 years before that in cyclic operation; over 100 gasifiers successfully operated outside the U.S. Selected in 1977 for two DOE demonstration projects.
CONDITONS:	Pressure: Atmospheric. Temperature: Gasification Zone = 2200°F. Gas Temperature: 250°F top gas, 1200°F clear gas.
REACTANTS:	Sized coal (+1/4 to 1 in. or +1/2 to 1-1/2 in.) with free-swelling index less than 2-1/2; Steam (internally generated) = 0.25 lb per lb of coal; Air = 2.3 lb per lb of coal for air-blown operation.
PRODUCTS:	Low Btu gas (air blown), 175 to 205 Btu/scf. Medium Btu gas (oxygen blown), 280 Btu/scf. Medium Btu gas (cyclic), 330 Btu/scf.
FEED METHODS:	Storage and surge hoppers above gasifier; intermediate lock hopper.
ASH REMOVAL:	Ash removal by rotating grate, lock hoppers, or wet grate.
DESCRIPTION:	Turndown ratio is 100:25; vertical cylindrical construction with a rotating grate in the bottom of the reactor; can be started up in about 24 hours.

EXHIBIT 2-5

SUMMARY DESCRIPTION - WINKLER GASIFIER

NAME:	Winkler
DEVELOPER/ LICENSOR:	Davy Powergas, Inc. P.O. Box 36444 Houston, Texas 77036
TYPE:	Fluidized bed gasifier.
STATUS/ HISTORY:	This commercial process was developed in the late 1920s. Davy Powergas plans to test U.S. coals at 15 atm (210 psig) in a 10-TPD pilot plant. Most previous experience was with young, brown coals and their cokes in Germany, India and Turkey.
CONDITIONS:	

	Fluidzed Bed	Off-Gases
Temperature, °F	1800-2100	1700-2000
Pressure, psig	Atmospheric	Atmospheric

(Pressure operation under test; 4/1 turndown capability.)

REACTANTS:	Crushed coal (0 x 3/8 in.), steam, air (or oxygen).
PRODUCTS:	108,000 scf per ton of coal of low Btu gas (118 Btu scf). [62,000 scf per ton of coal of intermediate Btu gas (290 Btu/scf).]
FEED METHODS:	Screw feeder for noncaking coals; pretreatment of caking coals (free-swelling index greater than 2-1/2).
ASH REMOVAL:	Bottom ash removal by ash conveyor screw (70% of ash entrained in gas).
DESCRIPTION:	Vertical cylindrical construction; steel shell lined with refractory. Secondary injection of steam and air (or oxygen) above fluid bed completely gasifies entrained particles. If required, a radiant-heat boiler, in disengaging space, cools ash below softening temperature. Waste-heat train takes product gas and entrained ash concurrently down through steam superheater, steam generator, and air preheat. Entrained ash removed by settling and cyclones.

EXHIBIT 2-6

SUMMARY DESCRIPTION - KOPPERS-TOTZEK GASIFIER

NAME:	Koppers-Totzek
DEVELOPER:	Process was codeveloped by Heinrich Koppers, GmbH, (now Krupp-Koppers) of Essen, West Germany, and Koppers Company, Inc., of Pittsburgh, Pennsylvania.
LICENSOR:	Krupp-Koppers, GmbH Essen, West Germany
LICENSE U.S. and CANADA:	Koppers Company, Inc. Koppers Building Pittsburgh, Pa. 15219
TYPE:	Entrained flow slagging gasifier.
STATUS/ HISTORY:	Commercial process; thirty-nine installed units worldwide. Three new plants under construction. Pilot unit (36 TPD) operated in 1948 for U.S. Bureau of Mines at atmospheric pressure and oxygen-blown conditions.
CONDITIONS:	Combustion-Zone Temperature = 3500°F. Off-gas temperature ≃ 2700°F. Atmospheric pressure. (Pressurized units to be tested at 450 psig.)
REACTANTS:	Dried pulverized (70 to 90% -200 mesh) coal, oxygen, and steam. Process can handle caking coals as well as other solid carbonaceous or liquid fuels.
PRODUCTS:	50,000 to 78,000 scf (dry basis) per ton of coal feed of medium Btu (286 Btu/scf) gas. Gas yield depends on type of fuel or coal rank. No tars or condensible hydrocarbons are produced.
FEED METHODS:	Dry, pulverized coal fed by screw feeder to mixing nozzle, entrained in O_2 and steam; accepts all types of coals. Bituminous coals usually fed at 2% moisture, lignites at 8% moisture.
ASH REMOVAL:	Bottom ash to slag quench tank; entrained ash quenched in water spray. Cooled, granulated bottom ash removed through water.
DESCRIPTION:	Horizontal refractory-lined gasifier with two or four heads, each head containing two adjacent burners and each pair of heads forming an ellipsoid about the base of a vertical waste-heat boiler. Gasifier shell is steam jacketed and refractory lined.

SUMMARY DESCRIPTION - TEXACO GASIFIER

NAME:	Texaco Coal Gasification Process
DEVELOPER/ LICENSOR:	Texaco Development Corporation 135 East 42nd Street New York, NY 10017
TYPE:	Entrained flow gasifier.
STATUS/ HISTORY:	Commercially proven with liquid hydrocarbon feed-stocks; seventy plants. Pilot plant (15 TPD) tests coals at 350 psig, soon at 1200 psig. TVA to demonstrate 168-TPD unit.
CONDITIONS:	Slagging temperature (to ∿3000°F) in partial oxidation chamber; product gas (400 to 500°F) with direct quench; pressure is 300 to 1200 psig; 50% turndown possible. Operation at 1200 to 2500 psig proposed by W.R. Grace Company. Raw gas at gasifier operating temperature may be obtained by omitting direct quench.
REACTANTS:	Pulverized coal, water or steam and oxygen or air to partial oxidation chamber. Coal particle size is confidential; it has been reported variously as 70%-200 mesh or ≈0.1 mm diam.
PRODUCTS:	53,000 scf per ton of coal of medium Btu (253 Btu/scf) product gas.
FEED METHODS:	Preheated, coal-water slurry pumped to gasifier. Any coal can be handled.
ASH REMOVAL:	Quenched slag particles removed from gasifier in water slurry; ash contains <2% carbon. Alternate cooling method passes hot product gas through gas cooler where high-pressure steam is generated.
ESCRIPTION:	Vertical cylindrical pressure vessel (carbon steel) lined at upper end with refractories. Coal, steam and oxygen fed at top to react under slagging conditions. Product gas with entrained molten slag is quenched at bottom and slag is removed in slag quench bath. Slag discharged through slag pots while cooled product is cleaned in water scrubber.

EXHIBIT 2-8

TABLE 2-6

COMPARISON OF COMMERCIALLY AVAILABLE GASIFIERS
OPERATING CONDITIONS

Gasifier	Coal Size	Coal Feed Rate (tons/day/gasifier)	Pressure (psig)	Temperature (°F) Combustor	Temperature (°F) Gasifier	Temperature (°F) Exit
Fixed bed						
Lurgi, Dry Ash	1/8"-1 1/2"	400-600	350-450	1,800-2,500	1,100-1,500	700-1,100
Wellman-Galusha	1"-2"	80-120	atmospheric	2,400		1,000-1,200
Woodall-Duckham	1/4"-1 1/2"	200	atmospheric	2,200	1,200	
Fluidized bed						
Winkler	0-3/8"	700-1,000	atmospheric	1,800-2,100		1,700-2,000
Entrained beds						
Koppers-Totzek	pulverized	400	atmospheric	3,500		2,400
Texaco	pulverized	150-300	300-1,200	3,000		500

cycle: Lurgi and Texaco stand out in this regard. From the point of view of heat recovery and overall process efficiency, high temperatures are a decided advantage. We will see this clearly in the cycle studies.

The data in Table 2-7 compares the input requirements for both reactants and utilities. All systems appear competitive here.

Process outputs are shown in Table 2-8. Gas production differences are not significant. The steam production, however, should be noticed. This arises from the temperature of the gasifier process and will ultimately be reflected in the overall cycle performance since we require such steam for the cogeneration component of our system. The presence of tars and oils is an important consideration. Their effects become significant when we treat the environmental impacts of the system (see Section 4.0). The thermal efficiency differences show up in the coal use to provide given quantities of power and cogenerated steam. We quantify these effects in our discussion of the cycle, later.

Gas composition for typical coals (derived from data in the literature) is provided in Table 2-9. The effects of these differences are shown clearly in the cycle analyses later.

Finally, a brief summary comparing advantages and disadvantages for each process are provided in Table 2-10. Note that these are similar to the generic relations shown earlier in Table 2-4.

TABLE 2-7

COMPARISON OF COMMERCIALLY AVAILABLE GASIFIERS
INPUT REQUIREMENTS

Gasifiers	OXIDANT (lb/lb of coal) Air (low Btu gas)	OXIDANT (lb/lb of coal) Oxygen (medium Btu gas)	Steam (lb/lb of coal)	Water (gal/ton of coal)	Electric (kWh/ton of coal)
Fixed Bed					
Lurgi, Dry Ash	1.5-2.2	0.4-0.6	1.0-3.2	500-600	25
Wellman-Galusha	2.5-3.5	1.5	0.4-1.4	60	120
Woodall-Duckham	2.3	0.5	0.3-0.8	70	20-40
Fluidized Bed					
Winkler	1.7-3.0	0.4-0.7	0.4-0.7	200-300	10-25
Entrained Beds					
Koppers-Totzek		0.6-0.9	0.2-0.4	300-600	55
Texaco[1]		0.6-0.9	0.1-0.6	100-200	50

NOTES:
1. For slurry feed: slurry solids loading = 0.5-0.7.

TABLE 2-8

COMPARISON OF COMMERCIALLY AVAILABLE GASIFIERS
PROCESS OUTPUTS

Gasifiers	Low Btu Gas Production		Medium Btu Gas Production		Cold Gas Thermal Efficiency	BY-PRODUCTS		
							STEAM	
	Quantity (10^3 scf/ton)	Heating Value (Btu/scf)	Quantity (10^3 scf/ton)	Heating Value (Btu/scf)	(Coal to gas, %)	Tars and Oils (lb/ton of coal)	Low Pressure (lb/ton of coal)	High Pressure (lb/ton of coal)
Fixed Bed								
Lurgi, Dry Ash	100	179	60-70	280-300	63-80	100-140	3,500	1,300
Wellman-Galusha	130	120-170	60	270-290	69-76	120		
Woodall-Duckham	100	170-200	60	280-310	67-77	150	550	
Fluidized Bed								
Winkler	110	120	60	270-295	65-72	-		2,280
Entrained Beds								
Koppers-Totzek			50-80	270-290	75	-	1,000	3,000
Texaco			60	250-280	70-80	-		5,000

TABLE 2-9

COMPARISON OF COMMERCIALLY AVAILABLE GASIFIERS
SUMMARY OF GAS ANALYSIS

Input and Gas Composition	Lurgi (Fixed Bed)		Wellman-Galusha (Fixed Bed)	Woodall-Duckham (Fixed Bed)		Winkler (Fluidized Bed)	Kopper-Totzek (Entrained Bed)		Texaco (Entrained Bed)
Coal Type	Pittsburgh #6	Illinois #6	Pittsburgh #8	High Volatile Bituminous		Bituminous	Illinois #6	Eastern	Illinois #6
Oxidant	Oxygen	Air	Air	Oxygen	Air	Oxygen	Oxygen	Oxygen	Oxygen
Gas Composition (% by volume, dry after typical cleanup)									
CO	16.9	16.5	20.4	37.5	28.3	33.7	52.4	53.3	37.6
CO_2	31.5	13.4	8.7	18.0	4.5	16.3	10.0	10.0	20.8
CH_4	9.8	4.0	2.4	3.5	2.7	3.3	-	-	0.5
H_2	39.4	23.8	15.5	38.4	17.0	46.0	36.0	35.4	39.0
H_2S + COS	0.8	0.8	0.5	0.4	0.3	-	0.4	0.2	1.5
N_2 + Ar	1.6	41.5	52.5	2.2	47.2	0.7	1.1	1.0	0.6
Coal Heating Value (Btu/lb)	14,900	12,235	11,550	12,900	12,900	11,400	11,400	13,700	13,700
Product Gas Heating Value (Btu/lb)	285	302	140	280	175	290	290	286	286

TABLE 2-10

COMPARISON OF COMMERCIALLY AVAILABLE GASIFIERS

GASIFIERS	ADVANTAGES	DISADVANTAGES
Fixed Bed		
Lurgi, Dry Ash	Pressurized; turndown; caking coals.	Tars and oils; solids handling against high pressure.
Wellman-Galusha	Turndown; good efficiency.	Tars and oils; close bed temperature control required; low pressure.
Woodall-Duckham	Turndown; two stages; no direct water quenching required.	Tars and oils; limited to non-caking coals; bed temperature control required; low pressure.
Fluidized Bed		
Winkler	Turndown; all coals; clean effluent; low steam use.	Ash and char carryover; unconverted coal tram limits efficiency; low pressure.
Entrained Beds		
Koppers-Totzek	All coals; clean effluent; low steam use.	Small turndown; high oxygen requirement; ash removal problem; low pressure.
Texaco	All coals; clean effluent; turndown; pressurized.	Slurry feed; no real demonstration yet; close control of oxygen required.

2.4 REFERENCES

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2.2 Miller, C.L., "The U.S. Coal Gasification Program: Program and Projects", Mechanical Engineering, August 1980, 34-40.

2.3 Hartman, H.F., et al, "Low Btu Coal Gasification Processes, Volume 1, Summary Screening and Comparison", Oak Ridge National Laboratory, Oak Ridge, Tennessee, ORNL/ENG/TM-13/V1, November 1978.

2.4 Schora, F.C., and J.W. Loeding, "Low-Btu Gas as an Industrial Energy Source", presented at the Low-Btu Gas: Its Future Symposium, Dundee, Illinois, October 2-4, 1977.

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2.6 "Coal Gasification: The Koppers Process" Koppers Engineering and Construction, The Koppers Company, Pittsburgh, Pennsylvania, 1980.

2.7 Hartman, H.F., et al, "Low Btu Coal Gasification Process, Volume 2 Selected Process Description" Oak Ridge National Laboratory, Oak Ridge, Tennessee, ORNL/ENG/TM-13V2, November 1978.

2.8 Cavanaugh, E.C., et al, "Environmental Assessment Data Base for Low/Medium Btu Gasification Technology: Volume II, Appendix A-F" Industrial Environment Laboratory, U.S. Environmental Protection Agency, EPA-600/7-77-125b, November 1977.

3.0 COMBINED CYCLE PERFORMANCE

In this section we present cycle performance and optimization for an integrated combined cycle gasification plant operating in a cogeneration mode. Overall efficiency and performance are provided and compared for the commercially available gasifiers described in Section 2.0, integrated with a conventional combined cycle scheme.

As pointed out generally in Section 1.0, because of the coincident steam and electric demands of SPNC and the potential for significant energy savings, it is important to focus on the load matching and cogeneration impact of the combined cycle. Ultimately these considerations are reflected in the life cycle cost analysis (Section 8.0).

3.1 Combined Cycle Configuration

A schematic diagram of a generic integrated combined cycle/gasification plant plant is shown in Exhibit 3-1. A description of the major process steps including their auxiliary requirements and by-product follows:

- Oxygen Plant - Required for medium-Btu gas, the oxygen plant primarily consists of an air compressor, the air separation unit including heat exchangers, cold box components and expansion turbines, nitrogen compressors for purging requirements, and an oxygen compressor. Compressor drives may be either steam or electric and the expansion turbines may be designed to provide some of the remaining auxiliary requirements for the plant.

- Gasifier - The gasifier may be either oxygen or air blown and of the fixed, fluidized or entrained bed type. We note here that the amount of jacket steam produced by the gasifier may be significantly less than that

GENERIC COMBINED CYCLE/GASIFICATION PLANT FLOW DIAGRAM

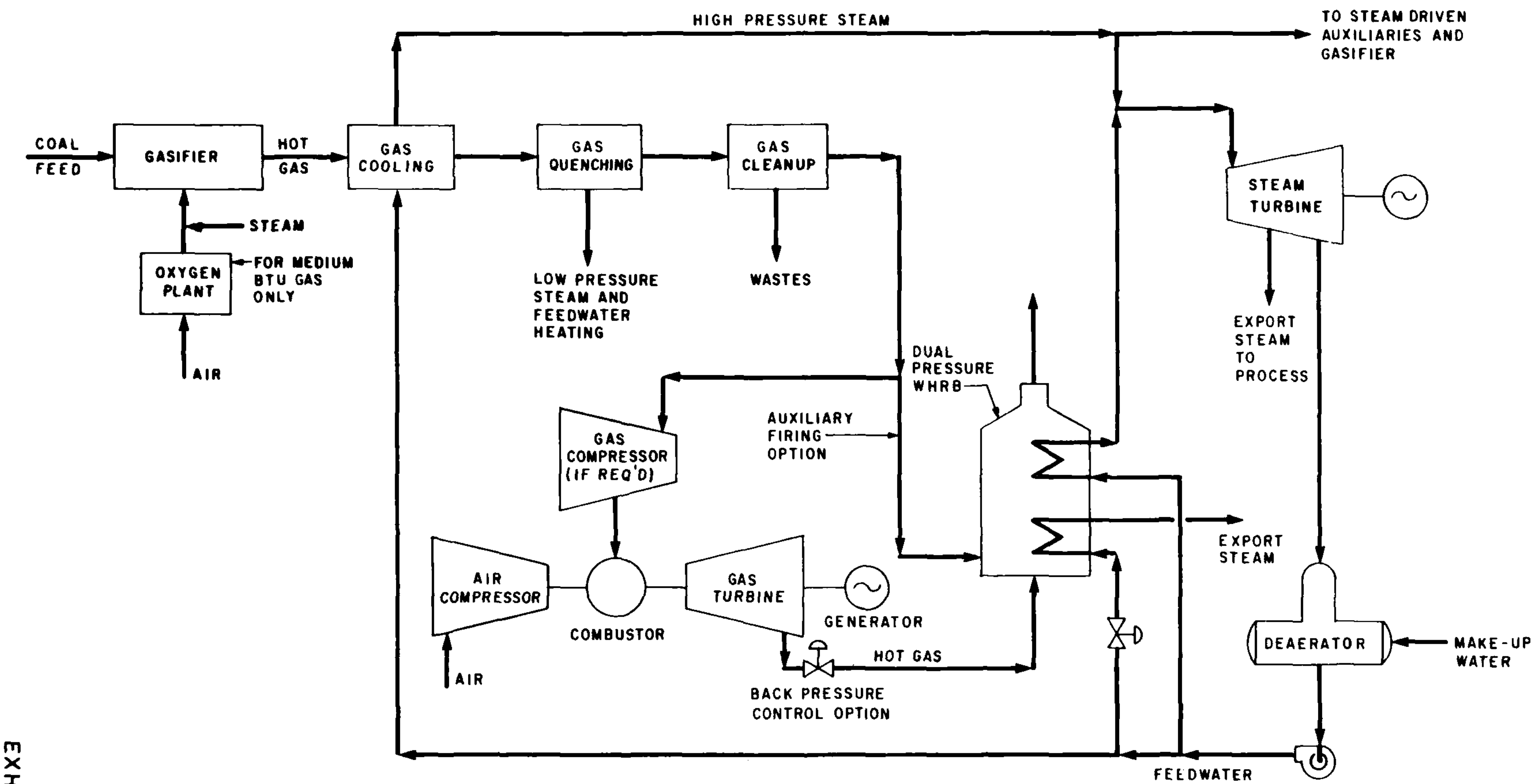


EXHIBIT 3-1

required for the gasification process. This is especially true for oxygen-blown fixed bed gasifiers, where large quantities of steam are required as a reactant moderator.

- Gas Cooling and Quenching - These processes are dependent on gasifier type and the method of gas cleanup; the details of their performance and overall input and output requirements are provided in Section 4.0. One observation is in order: in general, fixed bed single-stage gasifiers do not have sufficient sensible heat in the raw gas to produce high pressure steam. This is because of the lower gasifier exhaust temperature and tar liquor scrubbing that takes place before the gas enters a heat exchanger. Any additional heat recovered in the quenching phase may be used for producing low pressure steam and pre-heating feedwater make-up.

- Gas Cleanup - This process, which includes acid gas removal, is also described in detail in Section 4.0. Steam and electric auxiliary requirements are also given there. Following this step the gas is clean and cool.

- Gas Compression - This step is necessary for atmospheric gasifiers. The Lurgi and Texaco gasifiers do not require this step. However, for those, oxygen compression prior to the gasifier is required to a higher pressure level than that of the gas turbine combustor. The gas compressor may either be on the same shaft as the gas turbine or it may be steam driven.

- Gas Turbine Generator - Except for the combustor the gas turbine generator component is the same as one used in conventional combined cycle plants. Combustor modification is necessry due to the nature of the synthetic gas.

- Coal derived gas is produced as either low (90-160 Btu/scf) or medium (200-300 Btu/scf) Btu gas, depending on whether the gasifier is air blown or oxygen blown. The major difference between coal- derived gas and natural gas is in the reduced volumetric heating value, the quanitty of inerts and the chemical composition of the combustibles - carbon monoxide and hydrogen rather than methane. These differences affect the combustion process and cycle efficiency, and are reflected in the following quantities:

  - Fuel Throughput
  - Flame Temperature
  - Reaction Rate
  - Water Vapor Content
  - Non-Gaseous Contaminents
  - Emission Yielding Compounds

- Information in the literature indicates that with combustor modifications medium-Btu coal derived gas can be fired in present-day gas turbine units (Reference 1). Indeed one major gas turbine manufacturer is now offering a new design of oil-or gas-fired packaged combined cycle plants with built-in provisions for later conversion to coal derived fuels including medium-Btu gas (see Reference 2).

- Waste Heat Recovery Boiler (WHRB) - The sensible heat in the turbine exhaust gas is recovered and converted to steam in this step.

- Steam Turbine - The steam from the WHRB and the gas cooling process, if any, is expanded in a conventional steam turbine. Steam is taken, as necessary, for process and auxiliary requirements. The steam turbine

may operate in either a condensing or back-pressure mode.

One final word on the importance of plant integration is in order. The scheme shown in Exhibit 3-1 presents the combined cycle as integrated with the gasification plant, i.e., there is an interchange of electric power, feedwater and steam between systems, in contrast to a non-integrated system where the gasification plant would produce its own steam requirements. Previous studies have shown an integrated system to be the most economical and one which results in the highest overall thermal efficiency (Reference 3). This is especially true for a cogeneration facility where all steam driven auxiliaries including those in the oxygen plant can operate in a back-pressure mode thereby increasing overall efficiency.

3.2 Cycle Considerations and Options

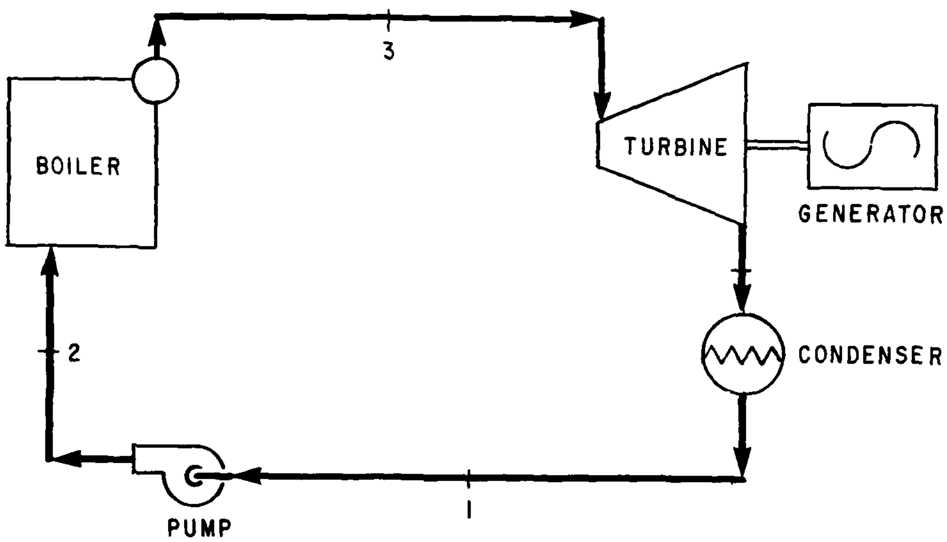
Before discussing the specifics of the cycle optimization and performance it is useful to review the two basic idealized thermodynamic cycles on which the gas and steam turbine cycles are based. These are the Brayton and Rankine cycles, respectively.

The cycle configuration and the temperature-entropy charts for these two cycles are shown in Exhibits 3-2 and 3-3. The cycle efficiency is given by:

$$\eta_{TH} = \frac{W_{net}}{Q_{Added}} = \frac{Q_{Added} - Q_{Rejected}}{Q_{Added}}$$

Where W represents the work done and Q the heat added or rejected. It is evident that the efficiency depends on the average temperatures at which heat is added and rejected. Any changes that maximize this difference lead to an increase in

RANKINE CYCLE CONFIGURATION & TEMPERATURE-ENTROPY CHART



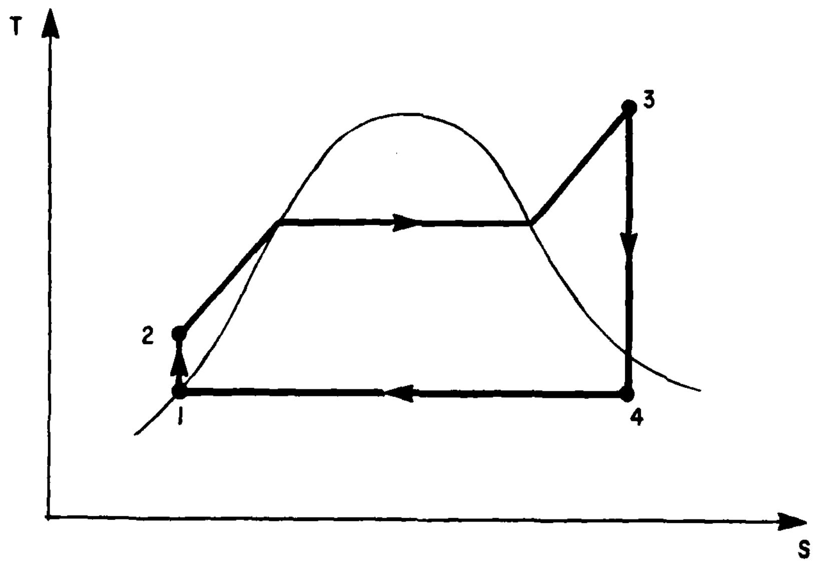
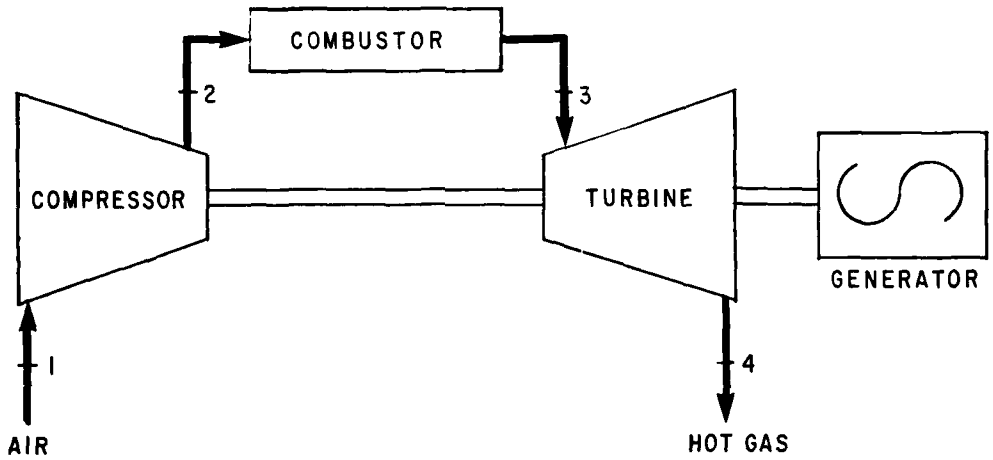


EXHIBIT 3-2

BRAYTON CYCLE CONFIGURATION & TEMPERATURE-ENTROPY CHART



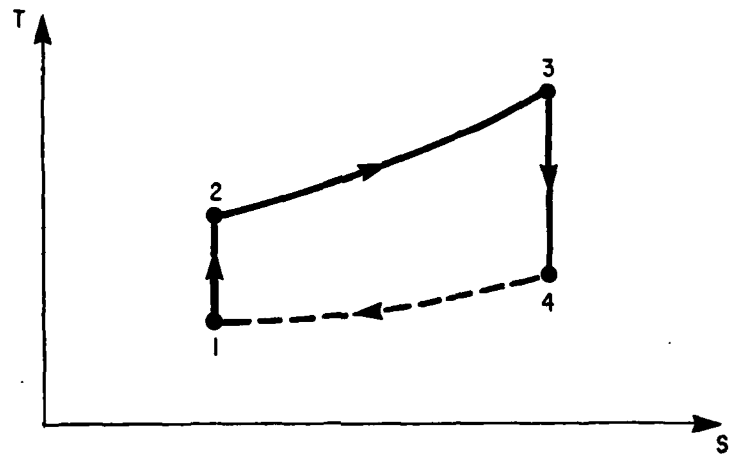


EXHIBIT 3-3

efficiency. This is accomplished in conventional steam plants by such cycle enhancements such as steam reheat, feedwater heating, increased boiler pressure and temperature, and reduced condenser pressure.

Because steam temperature is limited to approximately 1000°F by metallurgical considerations and gas turbine exhaust temperatures are in the range of 1000°F combining the two cycles, by topping the steam cycle with a gas cycle, will lead to an overall increase in cycle efficiency. The heat added to the cycle is kept at the higher temperature of the gas turbine and by recovering a large quantity of the heat in the Rankine cycle the overall heat rejected is reduced.

Another way of maximizing energy efficiency is through cogeneration. By reducing the amount of steam flow to the condenser the quantity of heat rejected is reduced, thereby leading to an increase in overall cycle efficiency. A combined cycle-cogeneration plant thereby makes use of both of these cycle enhancements.

A prime consideration in the assessment of any cogeneration scheme is its potential for matching thermal and electric loads, while, at the same time, remaining competitiv with power generated by a utility. This may require ryi.ıg the steam-to-power ratio generated by the prime mover. To ·crease this value below the baseline design point, a condensing, rather than back pressure steam turbine may be used. The ratio of throttle flow to condenser flow is varied to match loads. This arrangement has application in peak shaving schemes where a facility experiences sudden, sharp increases in load. This is not the case for the base load consideration of interest at SPNC. Increasing the steam-to-power ratio can be accomplished by either additional firing of coal derived gas in the waste heat recovery boiler or by increasing the gas turbine back pressure

thereby lowering the electric generation and increasing the amount of waste heat exhausted.

For the SPNC we consider the following base case cycle arrangement:

- Integrated combined cycle cogeneration with a back pressure steam turbine. For this case the amount of steam produced and electricity generated by the steam cycle is set by the gas turbine performance. This results in fixed thermal to electric ratio.

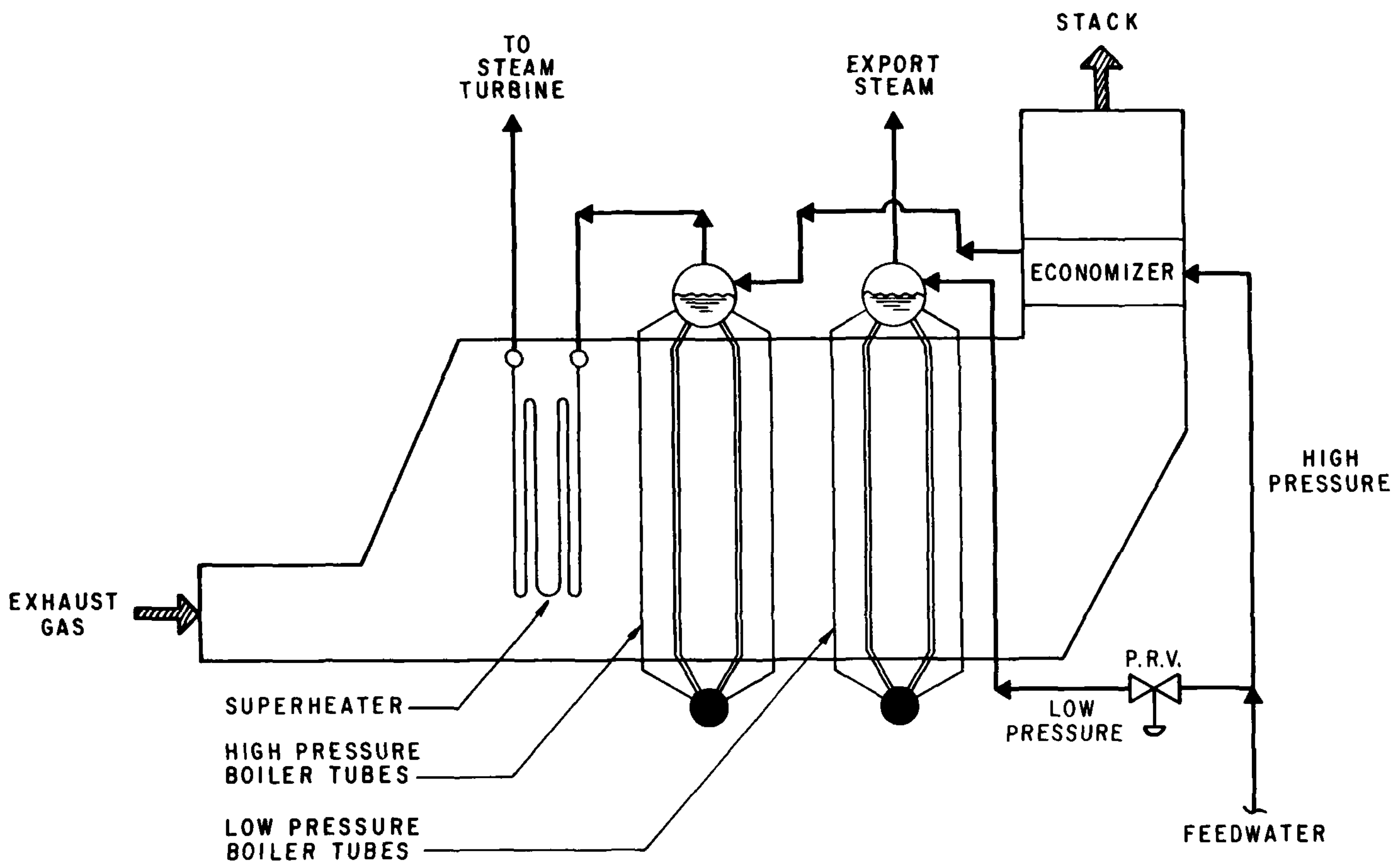
the following system components are investigated, as required, to ascertain their effect on plant efficiency:

- Low vs. Medium-Btu Gas
- WHRB Steam Pressure
- Dual Pressure WHRB (see Exhibit 3-4)
- Electric vs. Steam Driven Auxiliaries
- Steam vs. Gas Driven Gas Compressor
- Low Pressure Heat Utilization

Additional cycle enchancements including coal gas auxiliary firing, varying gas turbine back pressure and high temperature combustion are also studied.

It should be noted that because state-of-the-art combustor design precludes burning of low-Btu coal gas, the emphasis of cycle performance is on medium-Btu, i.e., oxygen blown, gasification. The air-blown gasifier cycle is presented only for comparison purposes or in one case, for the Wellman-Galusha gasifier, where adequate data on the oxygen-blown gasifier in unavailable.

DUAL PRESSURE BOILER



POPE, EVANS AND ROBBINS

EXHIBIT 3-4

3.3 Computer Model

The complexity of the integrated combined cycle requires a computer model be used to facilitate study of the numerous cycles in the performance evaluation. The computer model used has been developed by the Syntha Corporation, Greenwich, Connecticut, and is available on the Control Data Corporation Cybernet System. The program, which is an industry standard, can be used to determine heat balances for large scale nuclear and fossil fuel fired power plants.

The program can readily be applied to any configuration or arrangement of steam/water flow, heat transfer, gas flow, and/or mechanical components. It incorporates the ASME steam properties (1967) and the published procedures for prediction of steam turbine-generator performance. With the recent addition of gasifier components, the program embodies the most comprehensive library of components (i.e., technical content) available for heat and material balance.

The program utilizes "building-blocks" or elementary components, to model the physical components of a power plant as a schematic diagram easily translates into a standard input format. The program is then used to operate the model under various design options or under proposed plant performance conditions.

The Syntha Component Library (Reference 4) consists of four sections as follows:

- Steam and Water Flow Components - including steam turbines, pumps, pipes, valves, and other components necessary for flow stream modeling.

- Gas Flow Components - including gas turbines, compressors (fans), combustors, valves, pipes, and other components necessary for flow stream modeling.

- Heat Transfer Components - including components which transfer heat between gas, steam, and water, such as superheaters, boilers, gasifiers, economizers, feedwater heaters, deaerators, condensers, and gas/gas heat exchangers.

- Mechanical and Control Components - consisting of controls, schedules, generators, motors, loads, and shaft components for making mechanical connections.

A description of these components and their input requirements are given in Appendix B-1. Because of a lack of experience at Syntha with the gasifier component, a special effort was taken to continuously monitor it. The gasifier component performance did not originally satisfy PER specifications. However, a number of changes, based on our recommendations, have improved it so that its accuracy is now established and output acceptable.

3.4 Cycle Performance

The purpose of the cycle evaluation is two-fold. The first is to determine the optimum cycle that will serve as the basis for life cycle cost evaluation, and the second is to compare the cycle performance of different gasifiers. The evaluation is carried out for the six commercially available gasifiers described in Section 2.0. These fall into three categories, depending on gasifier type:

- Fixed Bed - Lurgi, Woodall-Duckham, Wellman-Gallusha
- Fluidized Bed - Winkler
- Entrained Bed - Koppers, Texaco

Because of the expected general similarities in cycle performance between gasifiers of the same category, detailed computer cycle optimization runs were carried out for only one gasifier

in each category. This enables us to determine the optimum cycle per category. This optimized cycle then serves as a base case for the comparison between different gasifiers in the same category. Sample computer outputs for each gasifier are provided in Appendix B-2.

Control for the computer program is provided by specifying the gas turbine generator output. Load analysis (see Section 1.0) indicates optimum base loads in the range of 50-60 MW electric and 270,000-290,000 lb/hr steam (340 psig, sat.). Because of the high electric auxiliary requirement in the gasification plant and in order to facilitate control of the computer model, the gas turbine generator output is set at 60 MW and steam turbine is set to provide all auxiliary loads (either steam or motor driven).

Note that a single turbine component can be used to model all steam driven auxiliary and geneator drives operating between the same pressures. Only the gas compressor and the feed pump drives are automatically accounted for in the program by connecting them to a shaft component. In cases where the remaining steam turbine generator output is insufficient to satisfy the auxiliary load requirement, the gas turbine generator net output is then r uced.

A number of performance constants, shown in Table 3-1, have been built into the computer model. These values remain the same, regardless of gasifier make. Other gasifier properties are consistent with those values given in Table 2-7. The coal types and resulting dry gas compositions are taken from Table 2-9.

Results For Entrained Bed Gasifiers
Koppers-Totzek Gasifier

A block flow diagram for a Koppers-Totzek gasification plant is shown in Exhibit 3-5. The gasifier operates at low pressures and the process is characterized by high exhaust gas tempera-

TABLE 3-1

CYCLE MODEL PERFORMANCE CONSTANTS

Feedwater Make-Up Temperature	55°F
Ambient Air Temperature	70°F
Ambient Air Relative Humidity	80%
Air/Gas Compressor Efficiency	85%
Gas Turbine Efficiency	90%
Gas Turbine Pressure Ratio	11
Steam Turbine Efficiency	72%
Feed Pump Efficiency	85%
Heat Transfer Component Radiation Losses	3%
Boiler Pinch Point	25°F
Economizer Outlet Subcool Temperature	5°F
Deaerator Pressure	5 psig
Export Steam Pressure	340 psig
Gas Cleanup Temperature	80°F
Combustor Temperature	1985°F

KOPPERS-TOTZEK GASIFIER MODEL

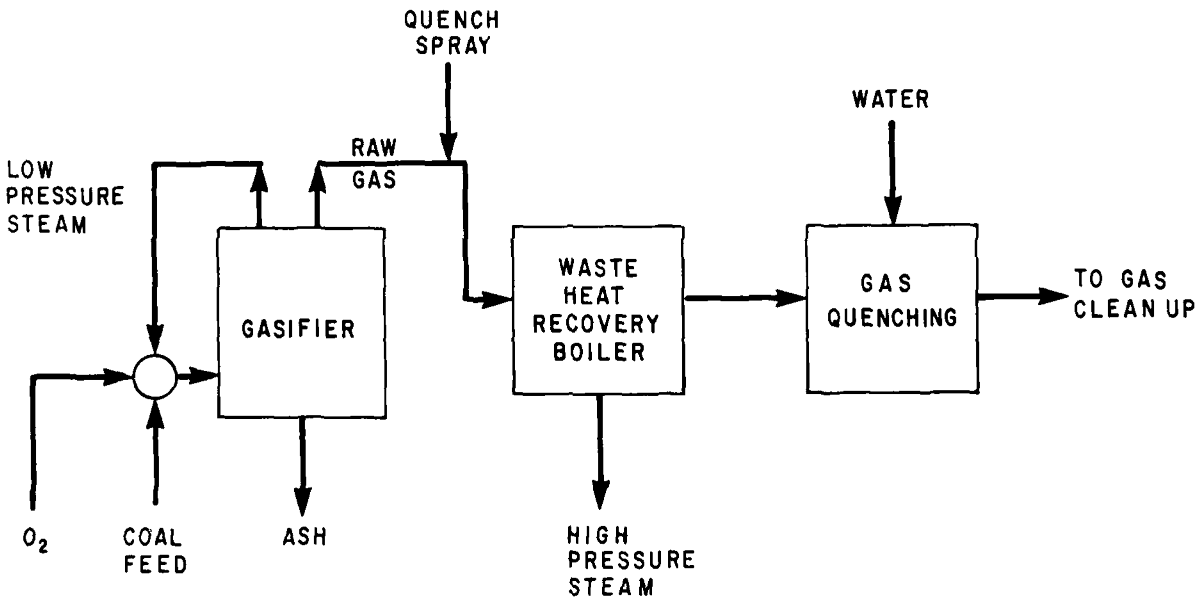


EXHIBIT 3-5

tures and a lack of tars and oils. This provides the opportunity for a high degree of heat recovery during the gas cooling process.

Gasifier properties were assumed as follows:

Pressure (psig)	Atmospheric
Exit Temperature (°F)	2700°F
Steam Use (lb/lb coal)	0.2
Oxygen Use (lb/lb coal)	0.7
Jacket Steam (lb/lb coal)	0.3
Cold Gas Conversion Efficiency	0.75

Note that although gasifier exit temperature is given as 2700°F, a spray process which serves to solidify ash carryover, reduces the gas temperature to approximately 2100°F before entering the gas cooling WHRB.

Auxiliary electric requirements for the plant are as follows:

Oxygen Plant	200 kW/T coal
Gasifier	55 kW/T coal
Balance to Plant	70 kW/T coal
Total Auxiliaries	325 kW/T coal

These figures are consistent with those given in the literature (Reference 3-5).

Four runs were made to establish the Koppers base case. These are summarized below:

1) Main steam pressure and temperature in both the gas cooling heat exchanger and the main WHRB is 885 psi and 800°F. The gas compressor is steam driven.

2) A second run uses steam pressure and temperature reduced to 585 psi and 700°F.

3) Next, the addition of a dual pressure boiler (885 psig, 800°F and 340 psig, sat.) is assessed.

4) Finally, the gas compressor is gas driven and throttle pressure increased to 885 psig.

A sample heat balance computer model diagram for the last cycle is shown in Exhibit 3-6. Table 3-2 summarizes the results for all runs. A heat and mass balance diagram is shown in Exhibit 3-7.

The following conclusions may be drawn from the results:

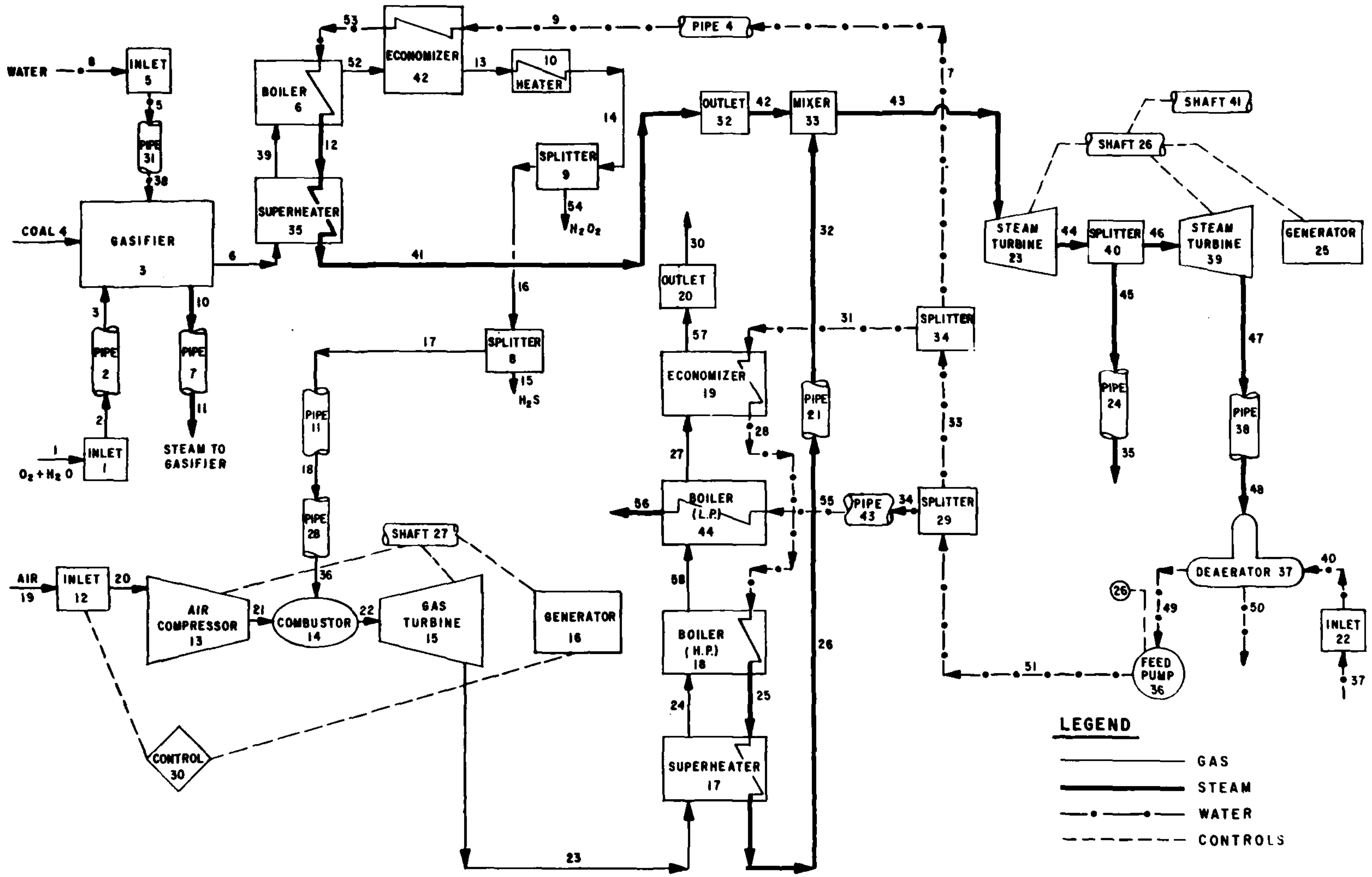
- Lowering throttle steam pressure and temperature alone does provide an increase in efficiency; however, a dual pressure boiler improves the cycle efficiency even more by increasing waste heat utilization in the WHRB.

- Although a gas turbine driven compressor does not have an advantage over a steam driven compressor in terms of efficiency, it does free the steam turbine capacity for other auxiliary needs.

Texaco Gasifier

A block flow diagram of a Texaco gasification plant is shown in Exhibit 3-8. It differs from the Koppers gasifier in that it operates at elevated pressures (up to 1200 psig). Another unique feature of the Texaco gasifier is that the coal is slurry fed. Water, rather than steam, is used as the reactant moderator. As with other entrained flow gasifiers, exhaust temperatures are high and the gas is essentially free of tars and oils, providing significant heat recovery in the gas cooling process.

HEAT BALANCE COMPUTER MODEL
KOPPERS-TOTZEK GASIFIER



POPE, EVANS AND ROBBINS

EXHIBIT 3-6

TABLE 3-2

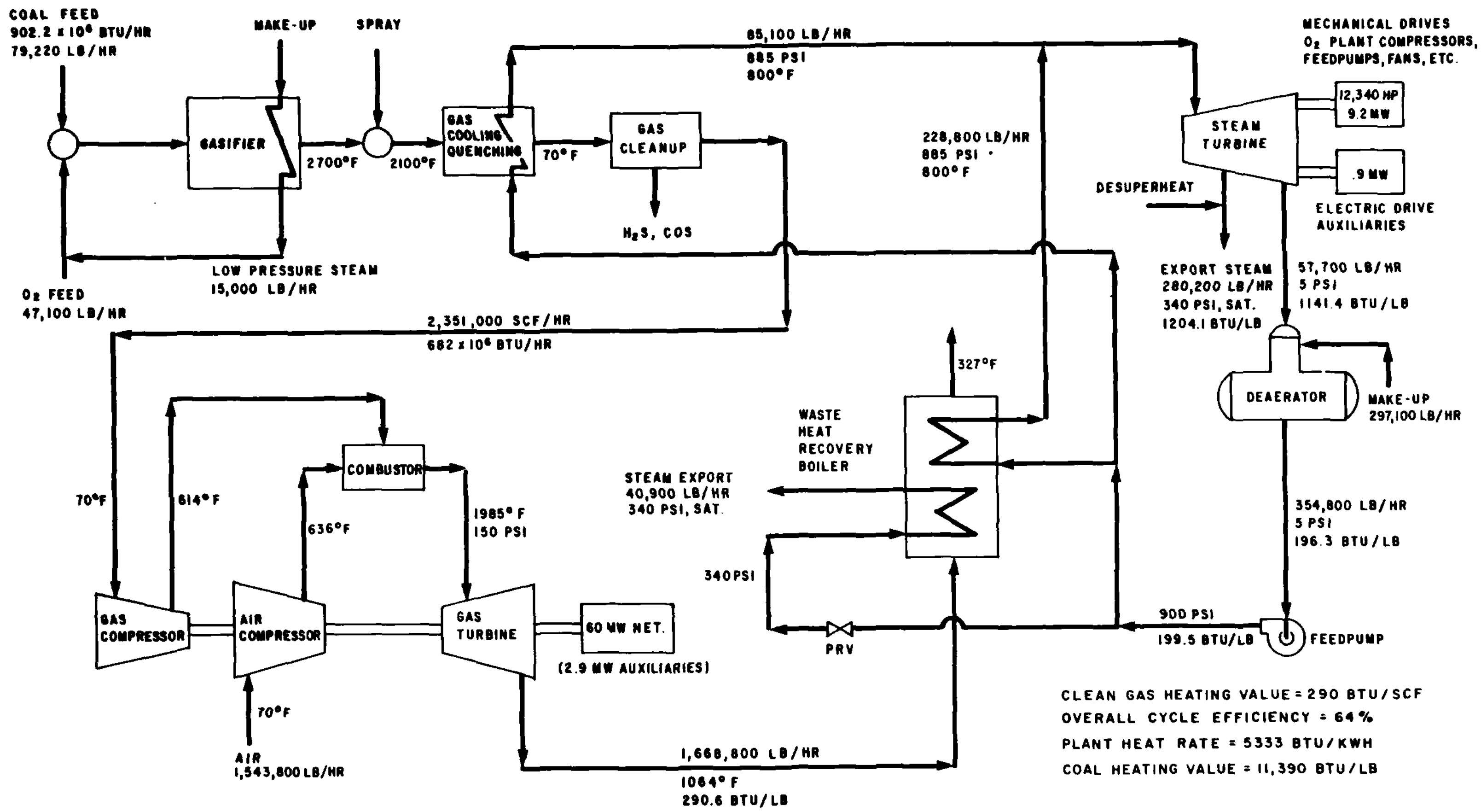
KOPPERS-TOTZEK INTEGRATED
COMBINED CYCLE PERFORMANCE

	Cycle 1	Cycle 2	Cycle 3	Cycle 4
Energy Input (10^6 Btu/hr)	804	804	804	902
Gasifier Oxident Feed (lb/hr)	42,000	42,000	42,000	47,100
Gasifier Steam Feed (lb/hr)	13,400	13,400	13,400	15,000
Gasifier Jacket Steam (lb/hr)	22,000	22,000	22,000	22,000
Clean Gas Flow Rate (10^3 scfd/hr)	2,098	2,098	2,098	2,350
Clean Gas Heating Value (Btu/scfd)	290	290	290	290
By-Product Tar and Oils (lb/hr)	-	-	-	-
Throttle Steam Pressure (psig)	885	585	885	885
Throttle Steam Temperature (°F)	800	700	800	800
Throttle Steam Flow (lb/hr)	298,600	318,500	278,400	313,900
Stack Outlet Temperature (°F)	361	335	327	327
Gas Turbine Output (MW)	60.0	60.0	60.0	60.0
Steam Turbine Output (MW)	3.0	0	2.7	10.1
Auxiliary Requirements (MW)	11.6	11.6	11.6	13.0
Gasifier Cold Gas Efficiency	0.75	0.75	0.75	0.75
Net Electric Generated (MW)	51.4	48.4	51.1	57.1
Net Export Steam Flow (lb/hr)	273,300	289,000	286,500	321,100
Thermal-to-Electric Ratio	1.88	2.11	1.98	1.98
Overall Efficiency	0.62	0.63	0.64	0.64

NOTES:

1. Includes both steam and electric driven auxiliaries, except for gas compressor and feed pump drives.

HEAT AND MASS BALANCE DIAGRAM
KOPPERS-TOTZEK GASIFIER

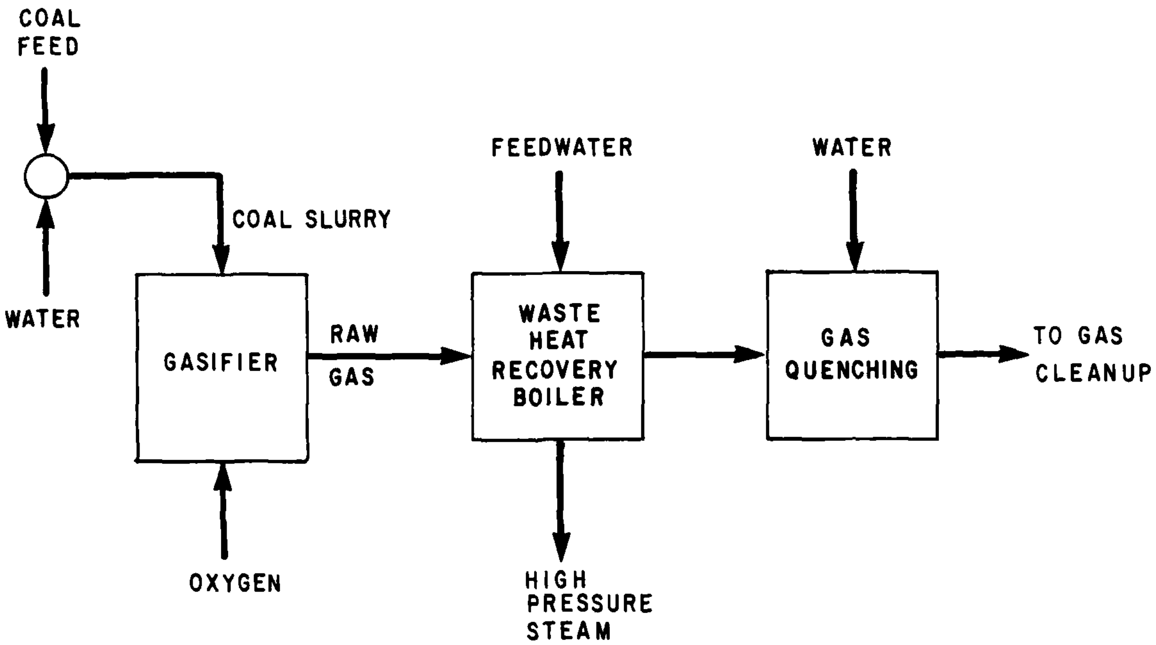


CLEAN GAS HEATING VALUE = 290 BTU/SCF
OVERALL CYCLE EFFICIENCY = 64%
PLANT HEAT RATE = 5333 BTU/KWH
COAL HEATING VALUE = 11,390 BTU/LB

POPE, EVANS AND ROBBINS

EXHIBIT 3-7

TEXACO GASIFIER MODEL



Gasifier properties are assumed as follows:

Pressure (psig)	585
Exit Temperature (°F)	2400
Slurry Solids Loading	0.67
Oxygen Use (lb/lb coal)	0.85
Cold Gas Conversion Efficiency	0.73

Note that the gasifier component in the computer model was designed to use steam and not water as the reaction moderator which is the case for the Texaco gasifier. To overcome this problem we artificially generate jacket steam in an amount equal to the liquid slurry flow thereby substituting steam as the moderator. The overall heat and mass balance for the gasifier is still maintained.

Auxiliary and electric requirements for the plant are as follows:

Oxygen Plant	320 kW/T coal
Gasifier	50 kW/T coal
Balance of Plant	84 kW/T coal
Total Auxiliaries	454 kW/T coal

Note that the oxygen plant auxiliaries includes the oxygen compressor.

One case was run for the Texaco gasifier corresponding to the optimized Koppers base case. The computer model flow diagram is shown in Exhibit 3-9. Results are summarized in Table 3-3. The cycle heat balance is shown in Exhibit 3-10.

Results For Fluidized Bed Gasifiers

Winkler Gasifier

A block flow diagram for the Winkler gasification plant is shown in Exhibit 3-11. The gasifier operates at low pressure. The

HEAT BALANCE COMPUTER MODEL
TEXACO GASIFIER

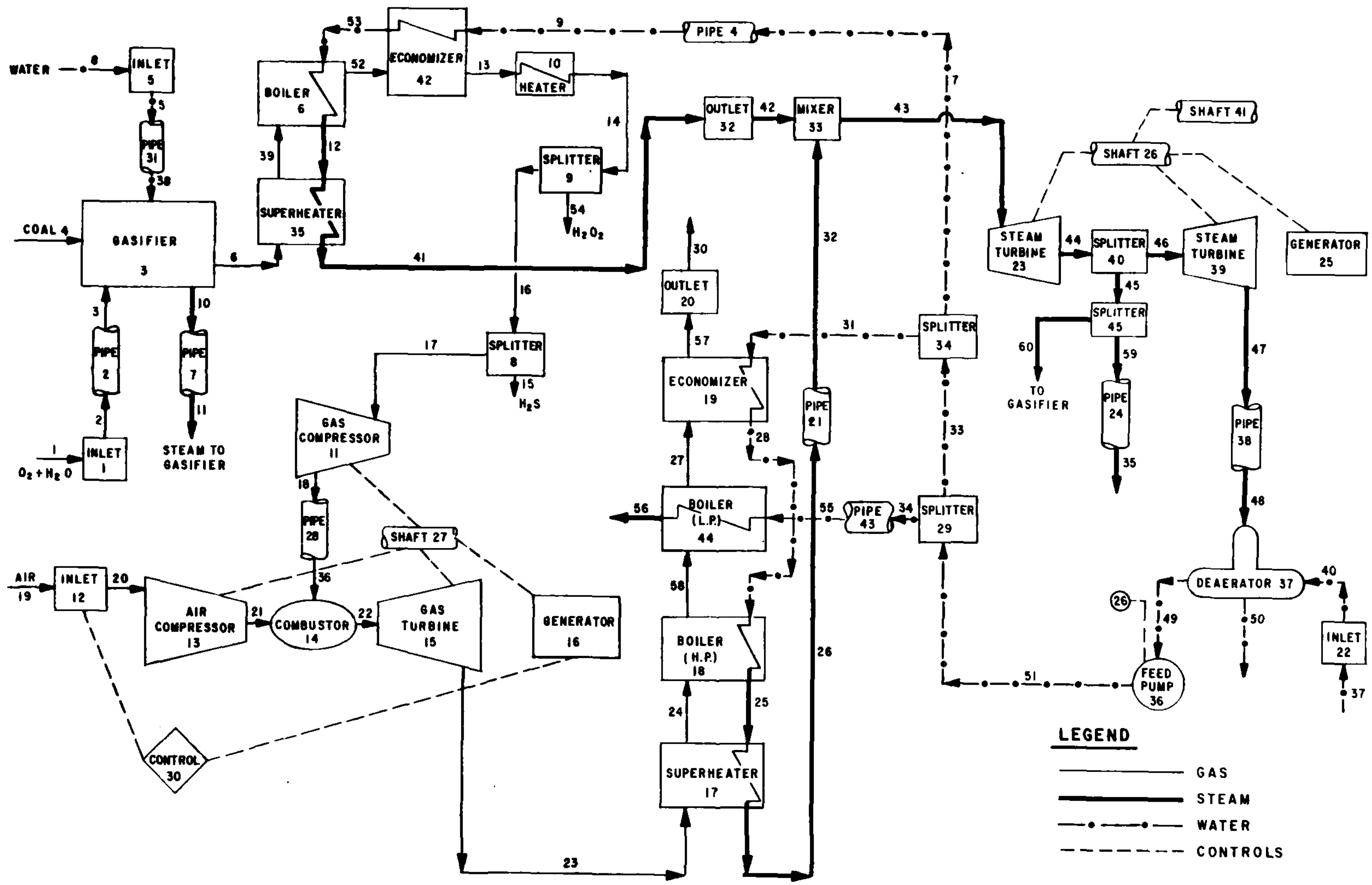


EXHIBIT 3-9

TABLE 3-3

TEXACO INTEGRATED
COMBINED CYCLE PERFORMANCE

Coal Feed (10^6 Btu/hr)	865
Gasifier Oxident Feed (lb/hr)	54,600
Gasifier Water Feed (lb/hr)	30,700
Gasifier Jacket Steam (lb/hr)	-
Clean Gas Flow Rate (10^3 scfd/hr)	3,346
Clean Gas Heating Value (Btu/scfd)	270
By-Product Tar and Oils (lb/hr)	-
Throttle Steam Pressure (psig)	885
Throttle Steam Temperature (°F)	800
Throttle Steam Flow (lb/hr)	328,000
Stack Temperature (°F)	325
Gas Turbine Ouput (MW)	60.0
Steam Turbine Output (MW)	10.5
Auxiliary Requirements (MW)[1]	14.3
Gasifier Cold Gas Efficiency	0.73
Net Electric Generated (MW)	56.2
Net Export Steam Flow (lb/hr)	330,000
Thermal-to-Electric Ratio	2.07
Overall Efficiency	0.68

NOTES:

1. Includes both steam and electric driven auxiliaries, except for feed pump drives.

HEAT AND MASS BALANCE DIAGRAM
TEXACO GASIFIER

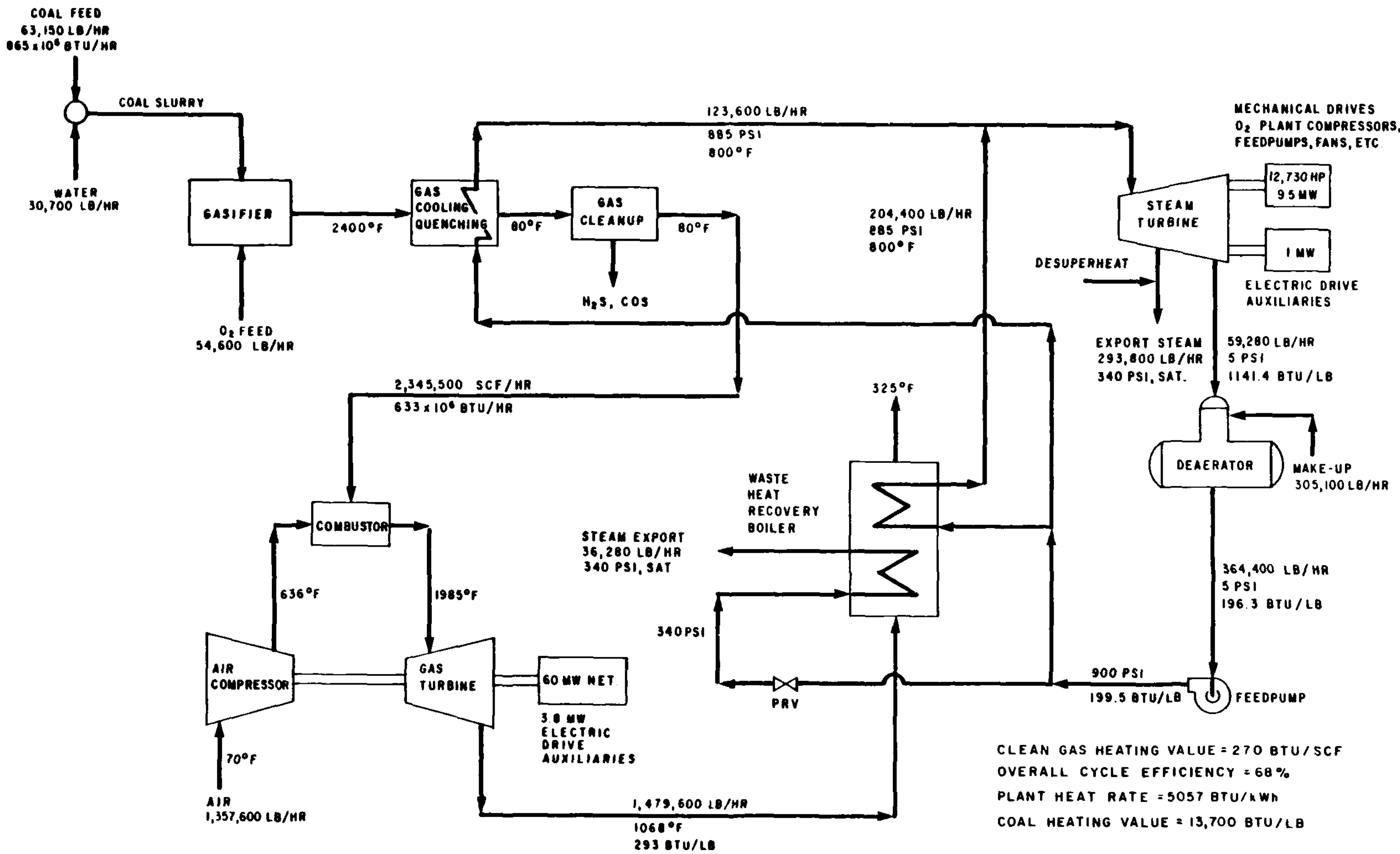
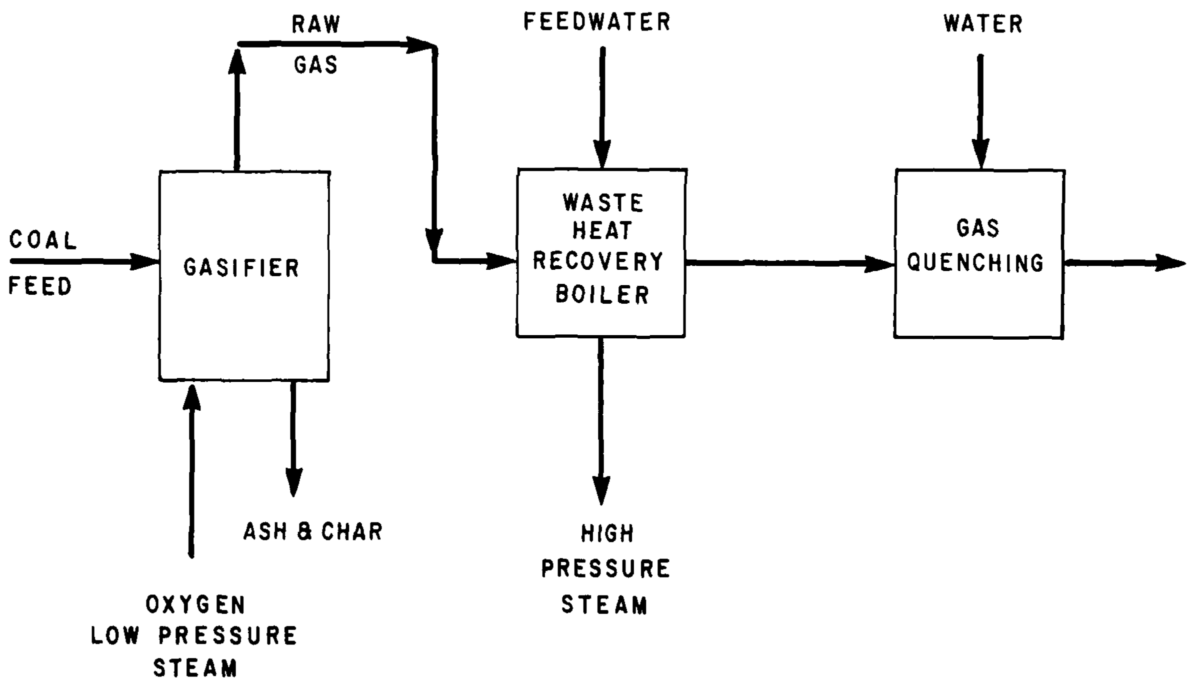


EXHIBIT 3-10

WINKLER GASIFIER MODEL



AD-A103 674 POPE EVANS AND ROBBINS NEW YORK F/G 10/2

EVALUATION OF COAL GASIFICATION/COMBINED CYCLE POWER PLANT FEAS--ETC(U)

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coal gasification reactions occur at relatively high temperatures, causing all tars and oils to be gasified. However, carbon conversion for the fluidized bed is lower than for the other gasifiers. Unreacted carbon leaves the gasifier as char, lowering the gasifier gas conversion efficiency.

Gasifier properties in the model were assumed as follows:

Pressure	Atmospheric
Exit Temperature	2100°F
Steam Use (lb/lb coal)	0.65
Oxygen Use (lb/lb coal)	0.65
Cold Gas Conversion Efficiency	0.66

Note that there is no generation of jacket steam. The relatively high quantity of steam required by the gasifier is obtained from the steam turbine extraction, lowering the amount of steam exported.

Auxiliary electric requirements for the plant are as follows:

Oxygen Plant	200 kW/ton
Gasifier	25 kW/ton
Balance of Plant	70 kW/ton
Total Auxiliaries	295 kW/ton

The four cases run for the Koppers gasifier were repeated for the fluidized bed Winkler gasifier. Results are summarized in Table 3-4. A sample computer model is shown in Exhibit 3-12. Note the additional stream splitting off the steam turbine extraction. A heat and mass balance diagram for the last cycle is shown in Exhibit 3-13.

The following observations can be made:

- As with the Koppers case the dual pressure boiler significantly increases the cycle efficiency and the gas

TABLE 3-4

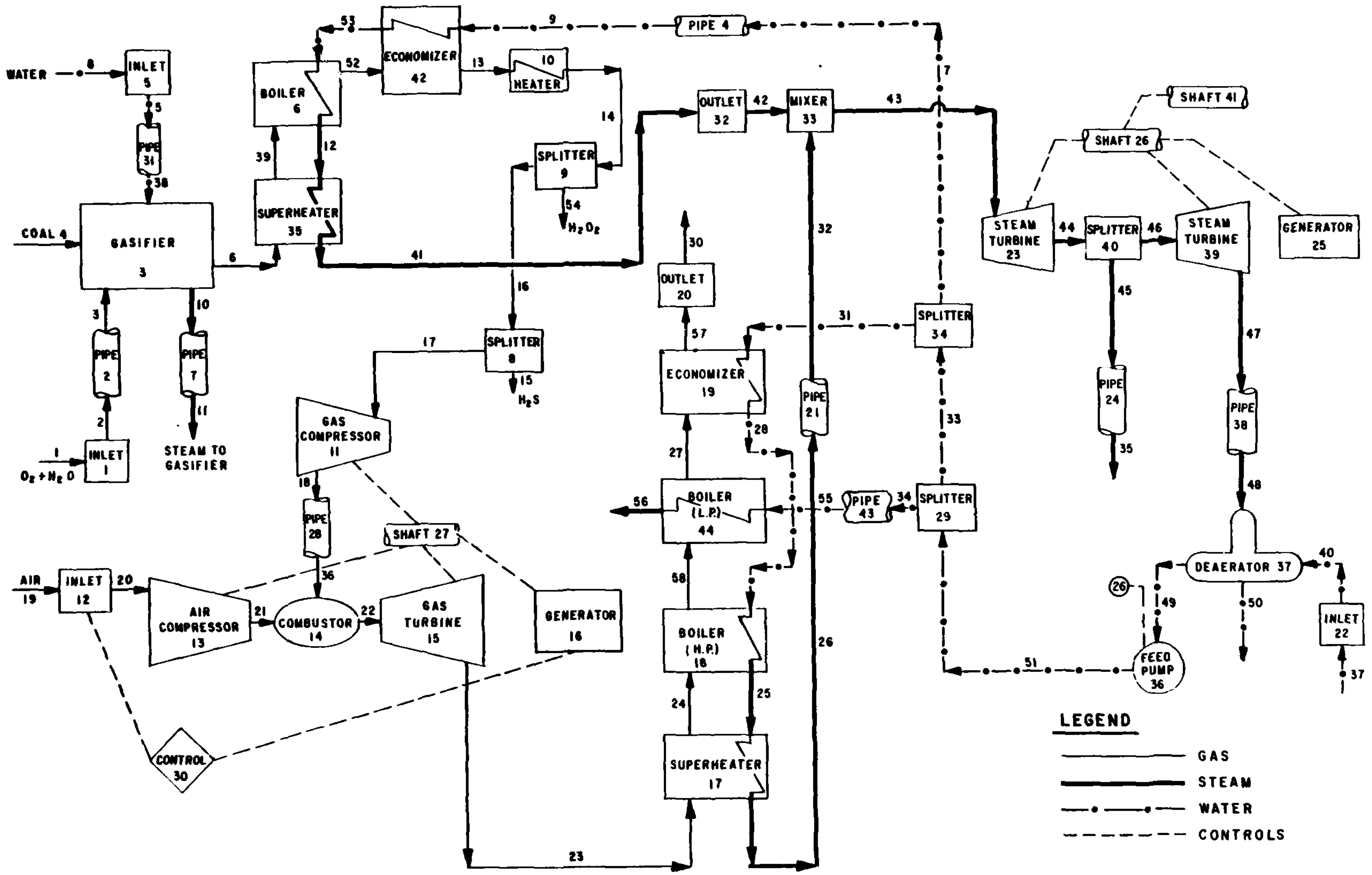
WINKLER INTEGRATED
COMBINED CYCLE PERFORMANCE

	Cycle A1	Cycle A2	Cycle A3	Cycle A4
Coal Feed (10^6 Btu/hr)	940	940	940	1,070
Gasifier Oxident Feed (lb/hr)	53,600	53,600	53,600	61,000
Gasifier Steam Feed (lb/hr)	53,600	53,600	53,600	61,000
Gasifier Jacket Steam (lb/hr)	0	0	0	0
Clean Gas Flow Rate (10^3 scfd/hr)	2,308	2,308	2,308	2,626
Clean Gas Heating Value (Btu/scfd)	270	270	270	270
Throttle Steam Pressure (psig)	885	585	885	885
Throttle Steam Temperature (°F)	800	700	800	800
Exit Gas Temperature (°F)	361	336	327	327
Throttle Steam Flow (lb/hr)	352,500	373,700	332,400	378,400
Gas Turbine Output (MW)	60.0	60.0	60.0	60.0
Steam Turbine Output (MW)	3.7	0	3.4	12.1
Auxiliary Requirements (MW)	12.0	12.0	12.0	13.7
Gasifier Cold Gas Efficiency	0.66	0.66	0.66	0.66
Net Electric Generated (MW)	57.7	48.0	51.4	58.4
Net Export Steam Flow (lb/hr)	268,000	273,000	285,800	325,300
Thermal-to-Electric Ratio	1.83	2.01	1.96	1.97
Overall Efficiency	0.53	0.52	0.55	0.55

NOTES:

1. Includes both steam and electric driven auxiliaries, except for gas compressor and feed pump drives.

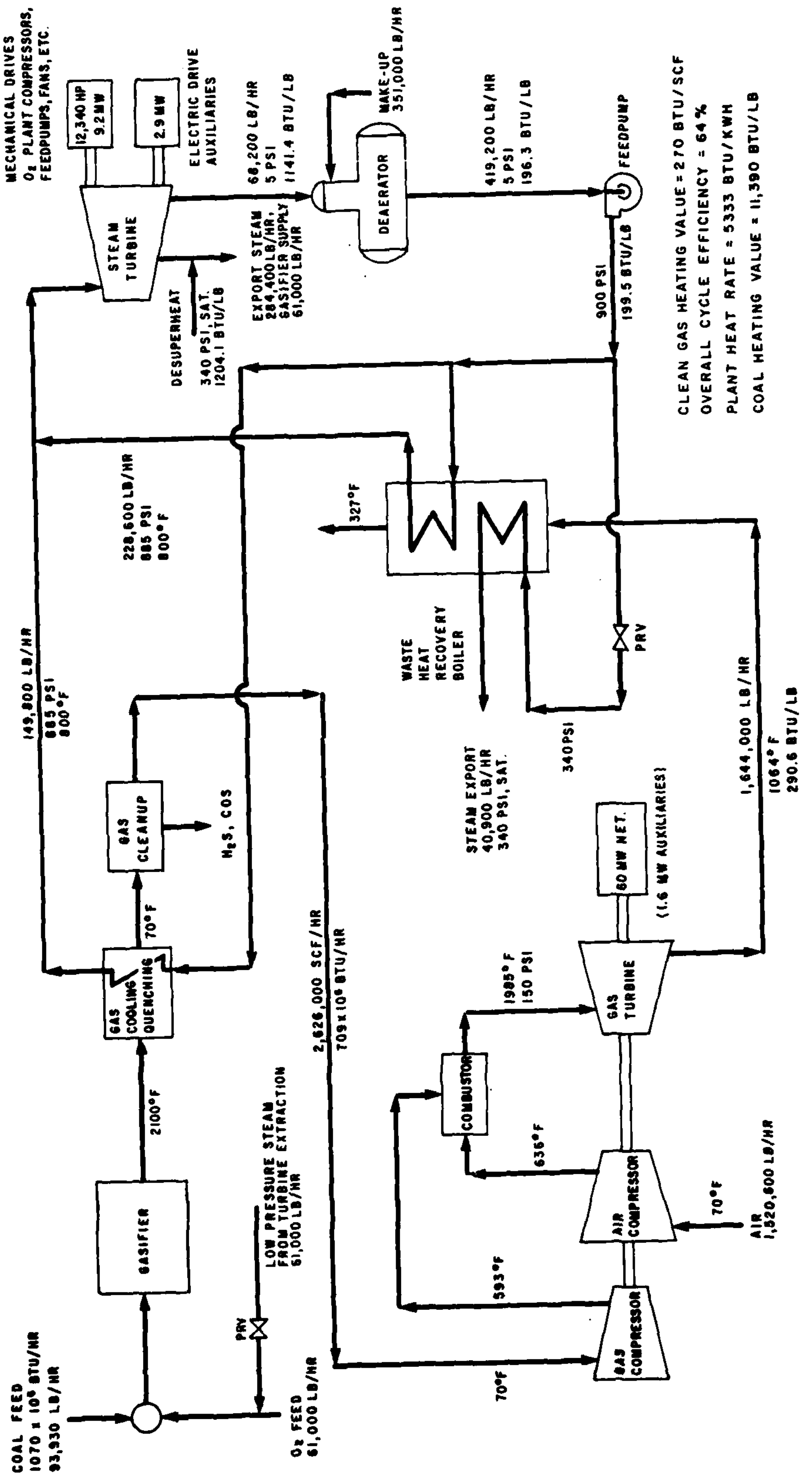
HEAT BALANCE COMPUTER MODEL
WINKLER GASIFIER



POPE, EVANS AND ROBBINS

EXHIBIT 3-12

HEAT AND MASS BALANCE DIAGRAM
WINKLER GASIFIER



POPE, EVANS AND ROBBINS

EXHIBIT 3-13

turbine driven gas compressor frees the steam turbine capacity for the other auxiliary requirements.

- The combination of char carryover which results in a lower cold gas efficiency and the high steam requirement as a gasifier reactant results in a lower efficiency for the Winkler gasifier cycle.

Results for Fixed Bed Gasifiers

Lurgi Gasifier

A block flow diagram of a Lurgi gasification plant is shown in Exhibit 3-14. The Lurgi gasifier operates under pressure (350-450 psi). As with all fixed bed gasifiers, tars and oils are formed and remain in the gas steam because of the relatively low gasifier temperature. The heavy tars must be scrubbed immediately following the gasifier exit. This, combined with the already low gas exit temperature, precludes any high pressure steam production during gas cooling.

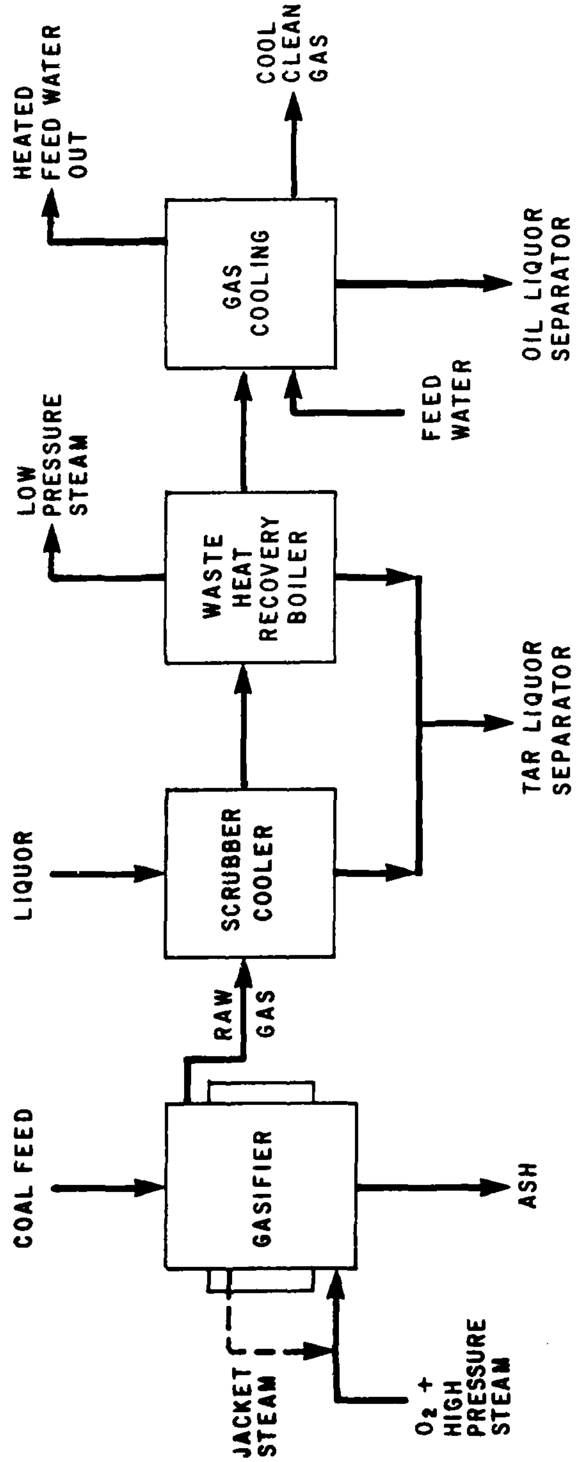
Gasifier properties are assumed as follows:

Pressure (psig)	345
Exit Temperature (°F)	700
Steam Use (lb/lb coal)	2.25
Oxygen Use (lb/lb coal)	0.4
Jacket Steam (lb/lb coal)	0.8
Tars and Oils (lb/lb coal)	0.07
Tars and Oils Heating Value (Btu/lb)	16,440
Cold Gas Conversion Efficiency	0.74

Note that although the gasifier exit temperature is given as 700°F, it is immediately reduced to 370°F following tar scrubbing.

Auxiliary electric requirements to the plant are as follows (Reference 3-5):

LURGI GASIFIER MODEL



POPE, EVANS AND ROBBINS

EXHIBIT 3-14

Oxygen Plant	90 kW/T coal
Oxygen Compressor	60 kW/T coal
Gasifier	25 kW/T coal
Balance of Plant	55 kW/T coal
Total Auxiliaries	230 kW/T coal

A sample heat balance computer model for the Lurgi base combined cycle plant is shown in Exhibit 3-15. Note the following:

- An additional boiler for supplemental tar and oil firing was added.
- Low pressure heat from gas cooling is used to peg the deaerator and to preheat feedwater make-up.
- Steam required as gasification reactant is greater than the amount produced in the gasifier jacket. This additional steam is taken from the steam exhaust.

The cases run for the Lurgi base case are summarized below:

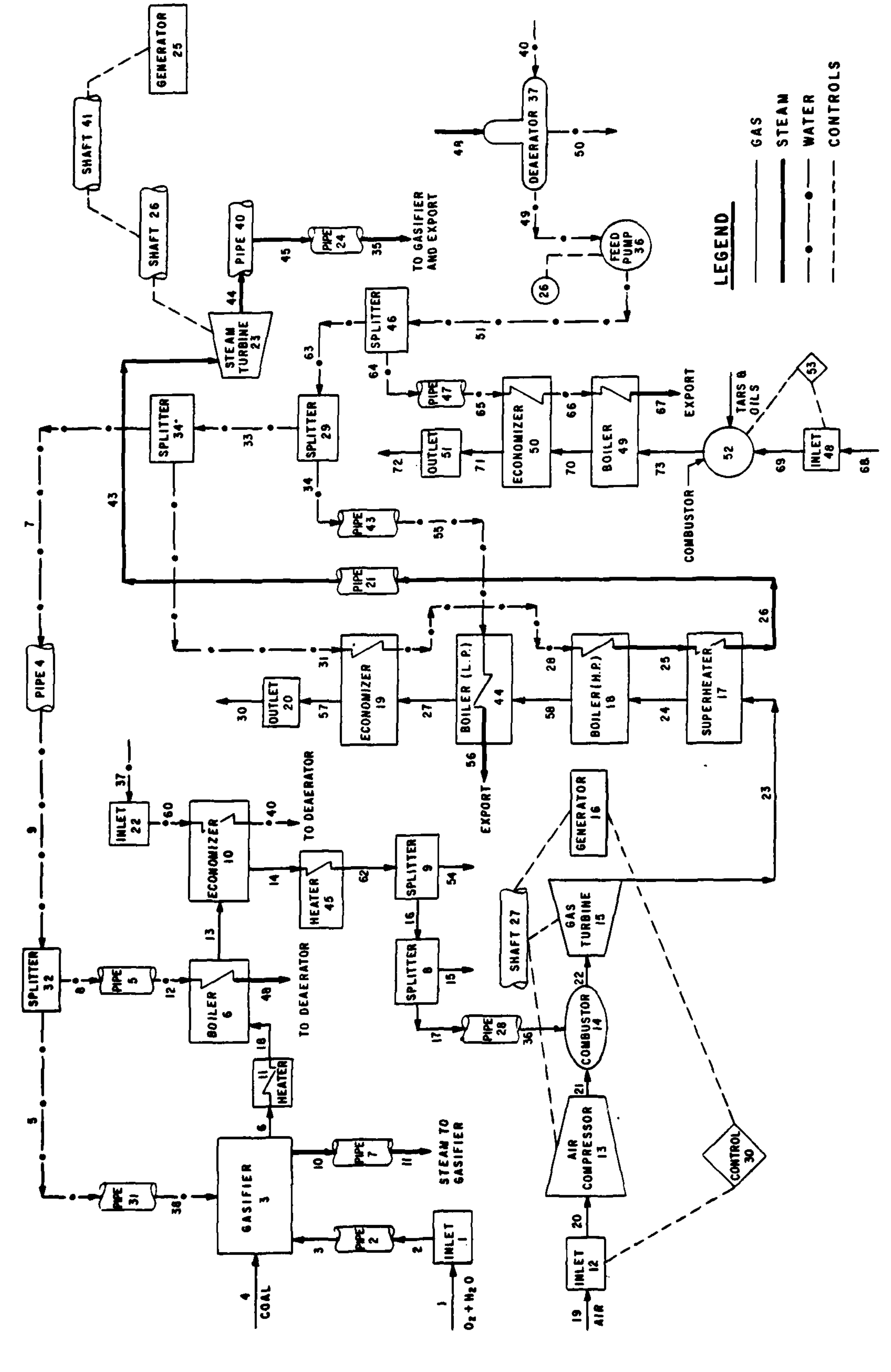
1) Main WHRB steam pressure and temperature is 885 psig and 800°F.

2) Next, pressure and temperature are reduced to 585 psig and 700°F.

3) Then, a low pressure boiler component is added to model a dual pressure WHRB. Steam throttle pressure was increased to 885 psig.

4) Finally, tar and oil by-products are combusted in a separate boiler (340 psig, sat.). Because stack clean-up for particulates is required for this boiler (see Section 4.0), the minimum stack gas temperature is set at 340°F.

HEAT BALANCE COMPUTER MODEL
LURGI GASIFIER



POPE, EVANS AND ROBBINS

EXHIBIT 3-15

The heat and mass balance diagram for the last cycle is shown in Exhibit 3-16. Results for all cases are summarized in Table 3-5. Several conclusions may be made from these results:

- The large quantity of high pressure steam required as a gasification reactant in addition to the lack of high pressure steam generation in the gas cooling process penalizes the overall thermal efficiency of the cycle.

- Lowering boiler pressure and temperature alone provide an increase in the overall efficiency, however, a dual pressure boiler has a higher efficiency.

- By-product tar and oil supplementary firing significantly increases the overall efficiency. However, the ratio of thermal to electric loads is still lower than the optimum of 290,000 lb/hr vs. 60 MW.

As a final example in the Lurgi based combined cycle, air-blown and oxygen-blown gasifers are compared in Table 3-6. The following observations are made:

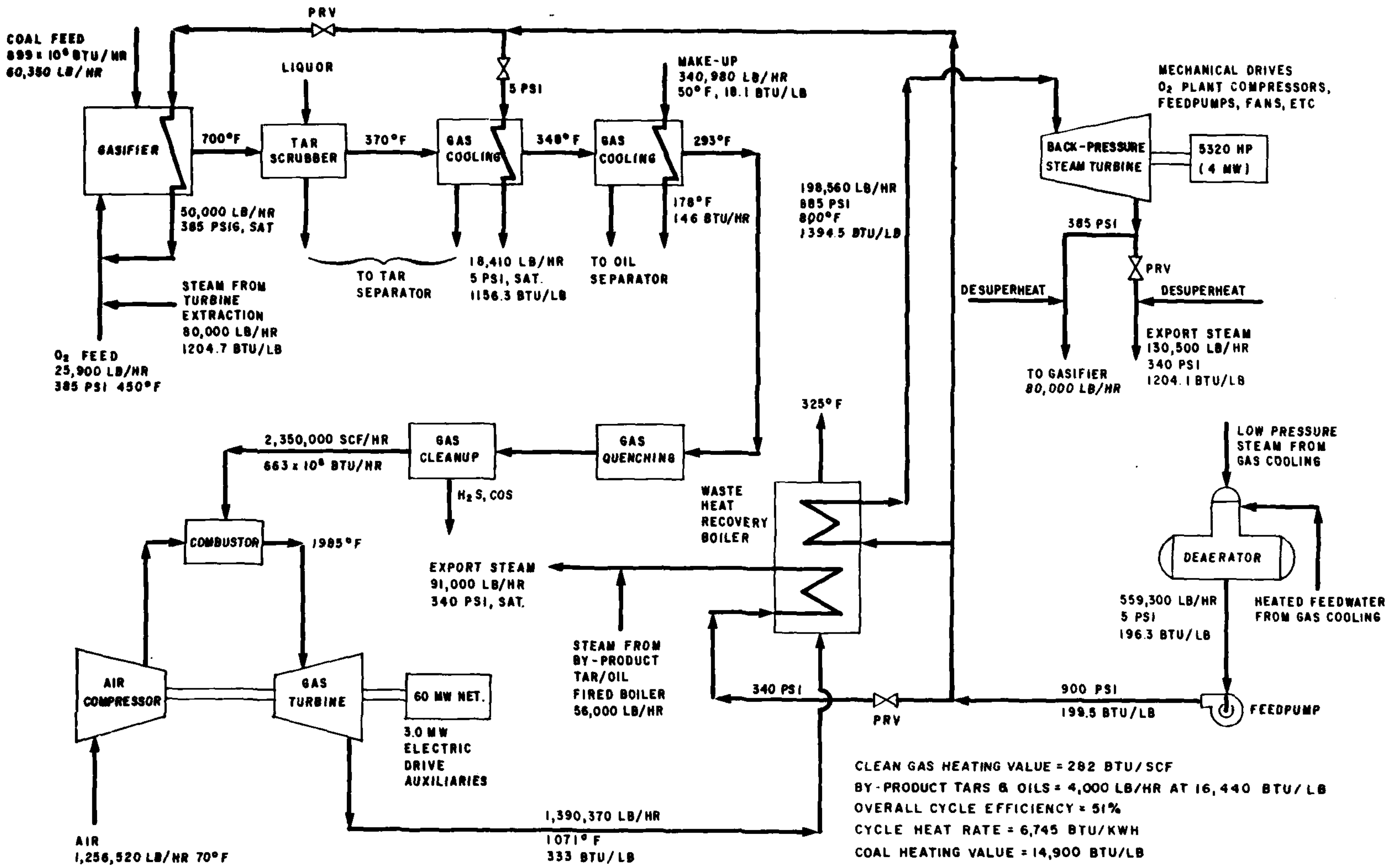
- The major difference between the two cases is in the quantity of steam required by the gasifier as a reaction moderator. More steam is required for the oxygen-blown case decreasing the net export steam flow.

- Though an oxygen plant is no longer required, the auxiliary power for air compression prior to gasification increase the total auxiliary requirements leading to a lower net electric generation.

Woodall-Duckham Gasifier

A block flow diagram of the Woodall-Duckham gasifier is shown in Exhibit 3-17. The gasifier operates at low pressure. The Woodall-Duckham differs from other fixed beds in that it is a

HEAT AND MASS BALANCE DIAGRAM
LURGI GASIFIER



CLEAN GAS HEATING VALUE = 282 BTU/SCF
BY-PRODUCT TARS & OILS = 4,000 LB/HR AT 16,440 BTU/LB
OVERALL CYCLE EFFICIENCY = 51%
CYCLE HEAT RATE = 6,745 BTU/KWH
COAL HEATING VALUE = 14,900 BTU/LB

POPE, EVANS AND ROBBINS

TABLE 3-5

LURGI INTEGRATED
COMBINED CYCLE PERFORMANCE

	Cycle 1	Cycle 2	Cycle 3	Cycle 4
Energy Input (10^6 Btu/hr)	899	899	899	899
Gasifier Oxident Feed (lb/hr)	25,900	25,900	25,900	25,900
Gasifier Steam Feed (lb/hr)	130,000	130,000	130,000	130,000
Gasifier Jacket Steam (lb/hr)	50,000	50,000	50,000	50,000
Clean Gas Flow Rate (10^3 scfd/hr)	2,350	2,350	2,350	2,350
Clean Gas Heating Value (Btu/scfd)	282	282	282	282
By-Product Tar and Oils (lb/hr)	4,000	4,000	4,000	4,000
Throttle Steam Pressure (psig)	885	585	885	885
Throttle Steam Temperature (°F)	700	700	800	800
Throttle Steam Flow (lb/hr)	218,200	233,700	198,600	198,600
Stack Temperature (°F)	360	337	326	326
Gas Turbine Output (MW)	60.0	60.0	60.0	60.0
Steam Turbine Output (MW)	4.4	2.2	4.0	4.0
Auxiliary Requirements (MW)	7.0	7.0	7.0	7.0
Gasifier Cold Gas Efficiency	0.74	0.74	0.74	0.74
Net Electric Generated (MW)	57.4	55.2	57.0	57.0
Net Export Steam Flow (lb/hr)	156,000	172,000	165,500	221,500
Thermal-to-Electric Ratio	0.96	1.10	1.02	1.37
Overall Efficiency	0.42	0.44	0.43	0.51

NOTES:

1. Includes both steam and electric driven auxiliaries, except for feed pump drive.
2. Includes by-products tar and oil fired boiler at 150 psig, sat.

TABLE 3-6

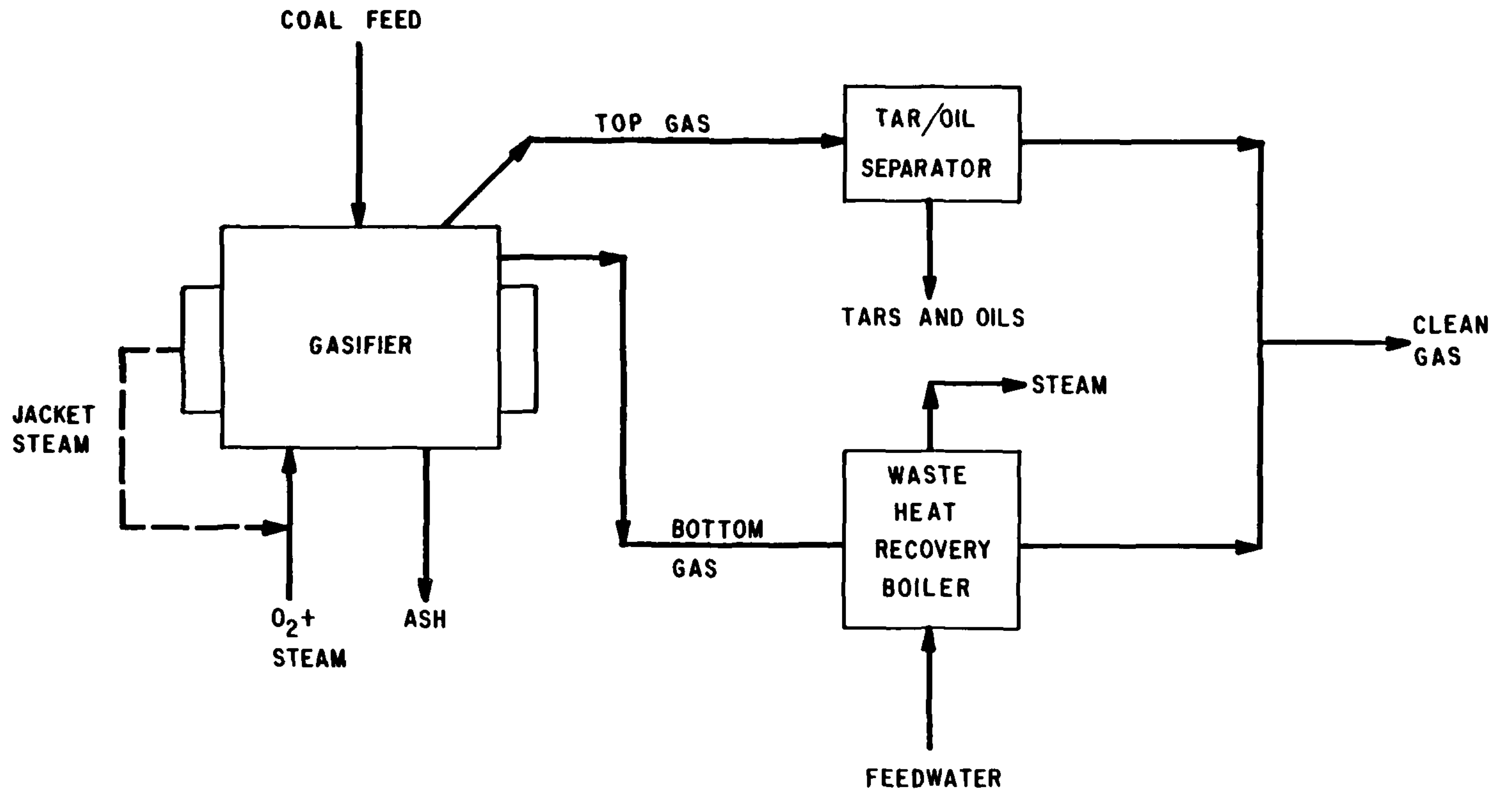
LURGI INTEGRATED
COMBINED CYCLE PERFORMANCE
AIR BLOWN VS. OXYGEN BLOWN GASIFIER

	Oxygen Blown	Air Blown
Energy Input (10^6 Btu/hr)	899	903
Gasifier Oxident Feed (lb/hr)	25,900	108,000
Gasifier Steam Feed (lb/hr)	130,000	70,000
Gasifier Jacket Steam (lb/hr)	50,000	50,000
Clean Gas Flow Rate (10^3 scfd/hr)	2,350	4,070
Clean Gas Heating Value (Btu/scfd)	282	166
By-Product Tar and Oils (lb/hr)	4,000	4,000
Throttle Steam Pressure (psig)	885	885
Throttle Steam Temperature (°F)	800	800
Throttle Steam Flow (lb/hr)	198,600	201,400
Gas Turbine Output (MW)	60.0	60.0
Steam Turbine Output (MW)	4.0	4.2
Auxiliary Requirements (MW)	7.0	12.0
Gasifier Cold Gas Efficiency	0.74	0.74
Net Electric Generated (MW)	57.0	52.2
Net Export Steam Flow (lb/hr)	221,500	337,000
Thermal-to-Electric Ratio	1.37	2.13
Overall Efficiency	0.51	0.64

NOTES:

1. Includes both steam and electric driven auxiliaries, except for feed pump drives.
2. Includes by-products tar and oil fired boiler at 150 psig, sat.

WOODALL-DUCKHAM GASIFIER MODEL



POPE, EVANS AND ROBBINS

EXHIBIT 3-17

two-stage gasifier. A portion of the gas is withdrawn from the vessel immediately following the gasification reaction. The remainder flows through a distillation retort section where it heats the descending raw coal. This gas (top gas) leaves the top of the gasifier at about 250°F. The bottom gas, which is essentially devoid of any tars and oils leaves at 1200°F. Because of this arrangement, only the top gas requires tar and oil removal. Bottom gas can directly enter a heat recovery boiler.

Gasifier properties were assumed as follows:

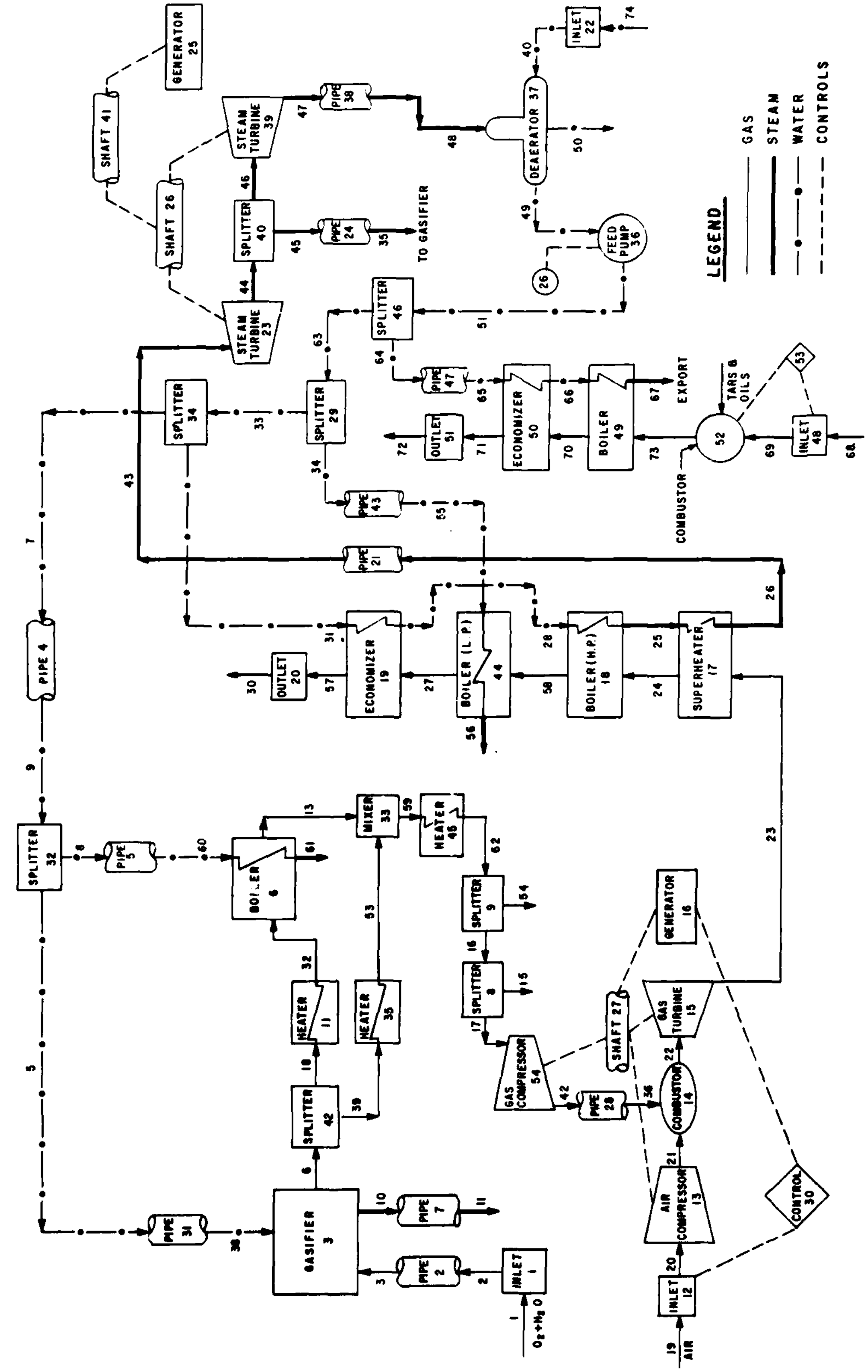
Pressure (psig)	Atmospheric
Top Gas Exit Temperature (°F)	1200
Bottom Gas Exit Temperature (°F)	250
Steam Use (lb/lb coal)	0.8
Oxygen Use (lb/lb coal)	0.5
Jacket Steam (lb/lb coal)	0.6
Tars and Oils (lb/lb coal)	0.05
Ratio of Top/Bottom Gases	1.0
Cold Gas Conversion Efficiency	0.74

Auxiliary power requirements are as follows:

Oxygen Plant	125 kW/ton
Gasifier	20 kW/ton
Balance of Plant	65 kW/ton
Total Auxiliaries	210 kW/ton

The heat balance computer model for the Woodall-Duckham gasifier is shown in Exhibit 3-18. Note that the gasifier component can not model a two-stage gasifier. This problem was overcome by specifying the exit gas temperature at the average of the two streams and then artificially splitting them into two, then heating one and cooling the other using heater components. The

HEAT BALANCE COMPUTER MODEL
WOODALL-DUCKHAM GASIFIER



POPE, EVANS AND ROBBINS

EXHIBIT 3-18

cycle performance is summarized in Table 3-7 and a heat and mass balance diagram is shown in Exhibit 3-19.

Wellman-Galusha Gasifier

A block flow diagram of the Wellman-Galusha gasification plant is shown in Exhibit 3-20. Due to the lack of experience and operating data for the oxygen-blown Wellman-Galusha gasifier, we present the performance of an air-blown plant noting that the net export steam, thermal-to-electric ratio and overall efficiency of the oxygen-blown plant will be somewhat lower.

The gasifier operates at atmospheric pressures and as with other fixed bed gasifiers produces a high quantity of tars and oils that are carried along in the gas stream. These tars and oils are scrubbed immediately following the gasifier exit. This is followed by gas quenching. There is no waste heat recovery step.

Gasifier properties are assumed as follows:

Pressure (psig)	Atmospheric
Exit Temperature (°F)	1000
Steam Use (lb/lb coal)	0.5
Air Feed (lb/lb coal)	2.8
Jacket Steam (lb/lb coal)	0.4
Tars and Oils (lb/lb coal)	0.07
Tars and Oils Heating Value (Btu/lb)	16,440
Cold Gas Conversion Efficiency	0.72

Auxiliary electric requirements are as follows:

Gasifier	120 kW/ton
Balance of Plant	65 kW/ton
Total	185 kW/ton

TABLE 3-7

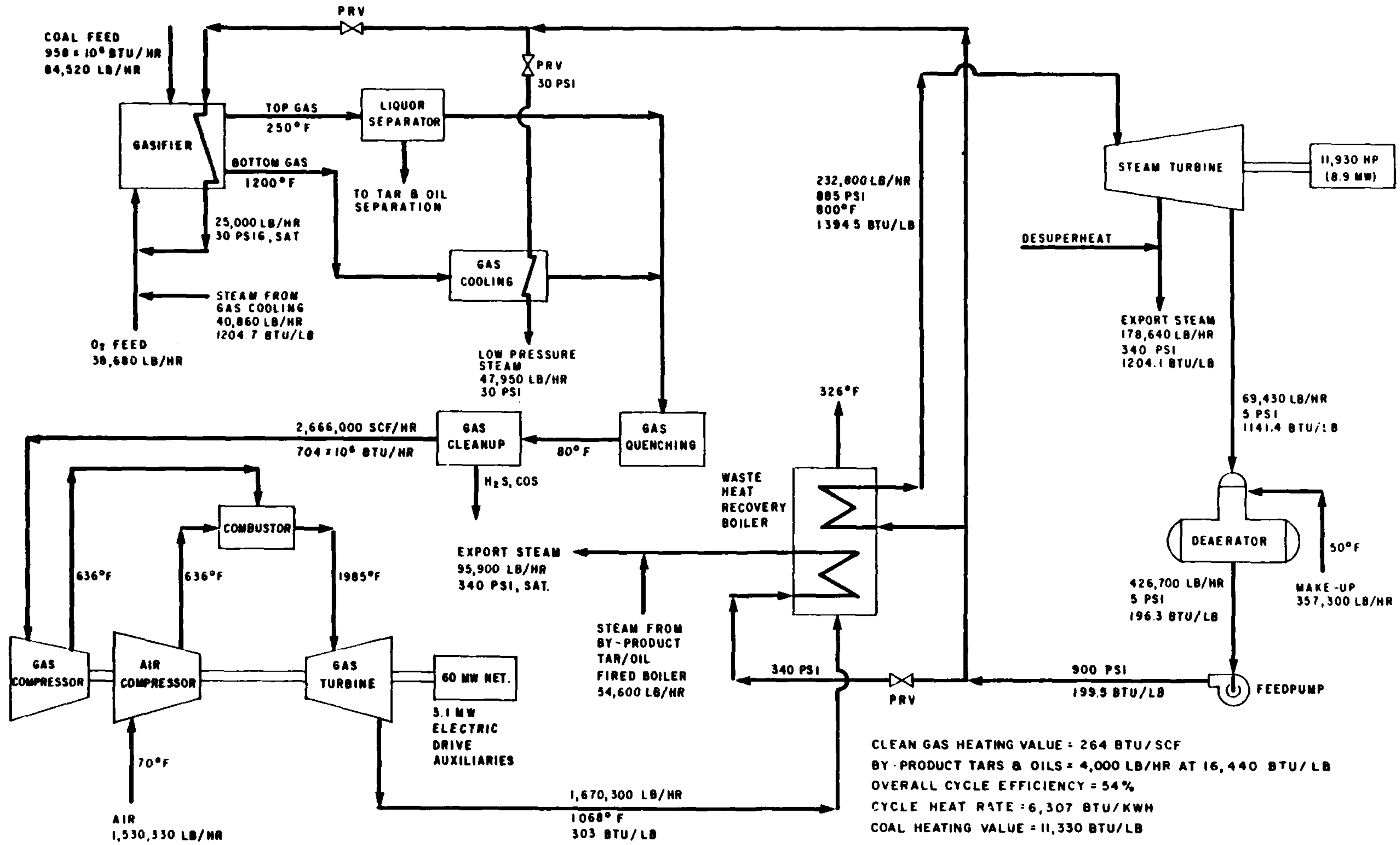
WOODALL-DUCKHAM INTEGRATED
COMBINED CYCLE PERFORMANCE

Energy Input (10^6 Btu/hr)	958
Gasifier Oxident Feed (lb/hr)	38,680
Gasifier Steam Feed (lb/hr)	65,860
Gasifier Jacket Steam (lb/hr)	50,000
Clean Gas Flow Rate (10^3 scfd/hr)	2,666
Clean Gas Heating Value (Btu/scfd)	264
By-Product Tar and Oils (lb/hr)	4,000
Throttle Steam Pressure (psig)	885
Throttle Steam Temperature (°F)	800
Throttle Steam Flow (lb/hr)	232,800
Stack Temperature (°F)	326
Gas Turbine Ouput (MW)	60.0
Steam Turbine Output (MW)	8.9
Auxiliary Requirements (MW)	12.0
Gasifier Cold Gas Efficiency	0.73
Net Electric Generated (MW)	56.9
Net Export Steam Flow (lb/hr)	274,600
Thermal-to-Electric Ratio	1.70
Overall Efficiency	0.54

NOTES:

1. Includes both steam and electric driven auxiliaries, except for feed pump drives.
2. Includes by-product tar and oil fired boiler at 340 psig, sat.

HEAT AND MASS BALANCE DIAGRAM
WOODALL-DUCKHAM GASIFIER



POPE, EVANS AND ROBBINS

EXHIBIT 3-19

WELLMAN-GALUSHA GASIFIER MODEL

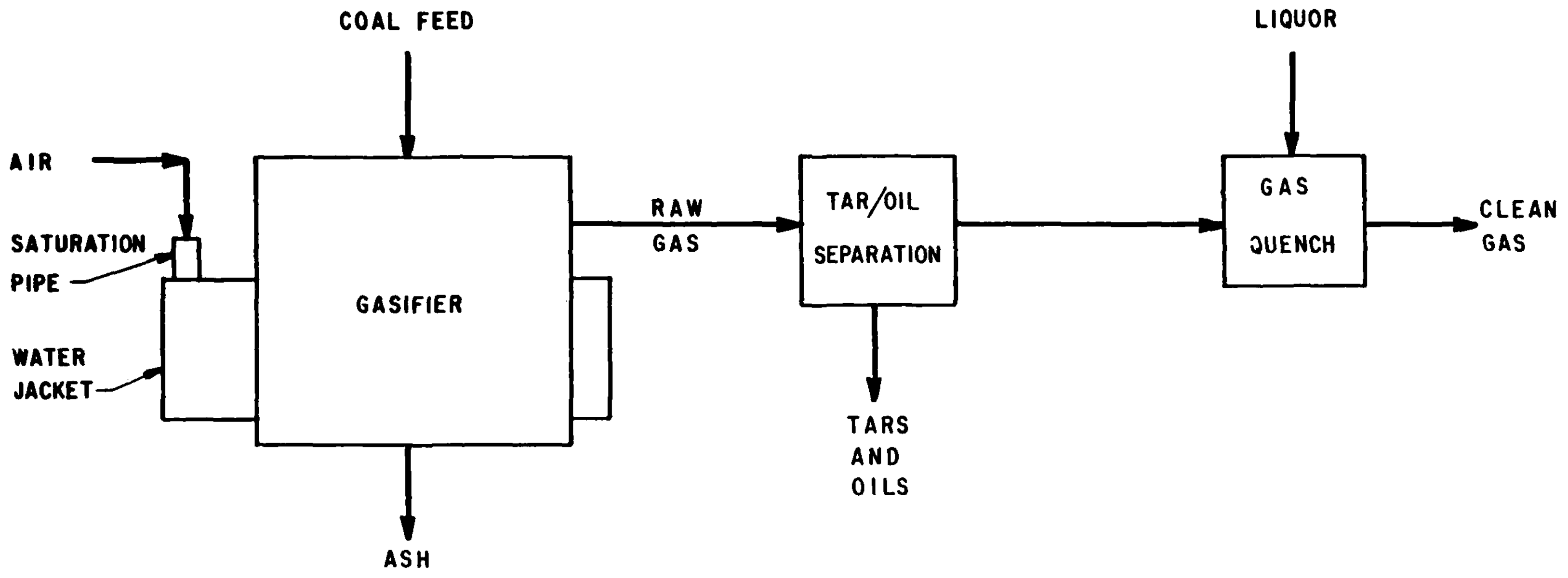


EXHIBIT 3-20

Note the relatively high auxiliary power required by the gasifier.

The heat balance computer model for the Wellman-Galusha gasifier corresponding to the optimized Lurgi cycle is shown in Exhibit 3-21. Results are summarized in Table 3-8 and a heat and mass balance diagram is shown in Exhibit 3-22.

3.5 Cycle Enhancements

In this section we build upon the base case cycle arrangements studied in the previous section. The following options are considered:

- Variable thermal to electric ratio.
- Increased combustion temperature.

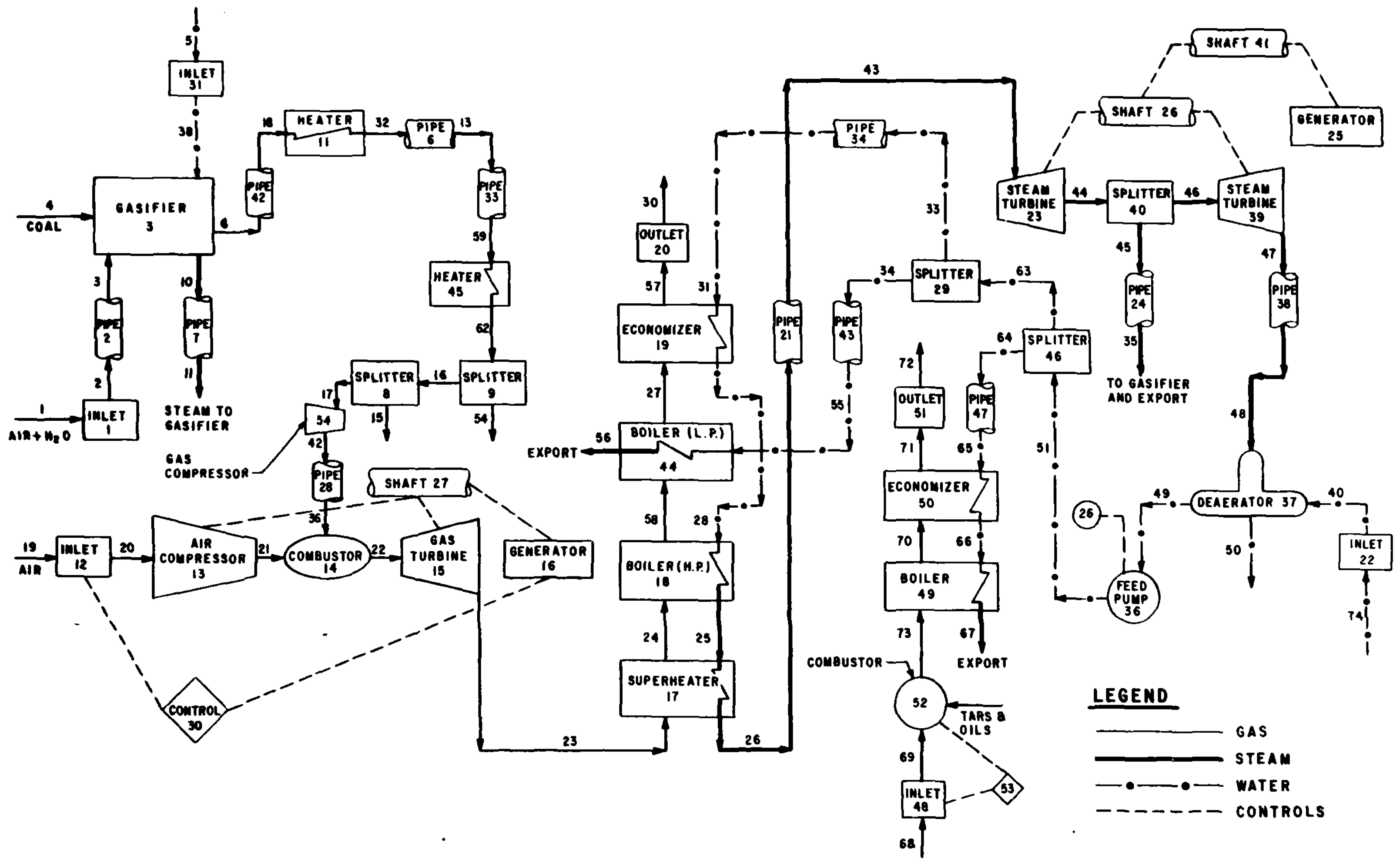
We note that combustion temperature is a function of the gas turbine generator component. The following configurations are considered to allow for a variation in thermal to electric ratio:

- Auxiliary coal-derived gas firing in the waste heat recovery boiler to increase the steam to electric ratio by adding fuel to the bottoming cycle.

- Varying the gas turbine exhaust back pressure, i.e., under-expanding, which increases the exhaust temperature thereby increasing the quantity of steam generated in the waste heat recovery boiler.

Both of the above increase the amount of steam generated - the first by burning more fuel; the second by reducing the electric generation.

HEAT BALANCE COMPUTER MODEL
WELLMAN-GALUSHA GASIFIER



POPE, EVANS AND ROBBINS

EXHIBIT 3-21

TABLE 3-8

WELLMAN-GALUSHA INTEGRATED
COMBINED CYCLE PERFORMANCE

Coal Feed (10^6 Btu/hr)	962
Gasifier Oxident Feed (lb/hr)	208,700
Gasifier Steam Feed (lb/hr)	36,800
Gasifier Jacket Steam (lb/hr)	30,000
Clean Gas Flow Rate (10^3 scfd/hr)	4,502
Clean Gas Heating Value (Btu/scfd)	154
By-Product Tar and Oils (lb/hr)	4,000
Throttle Steam Pressure (psig)	885
Throttle Steam Temperature (°F)	800
Throttle Steam Flow (lb/hr)	231,300
Stack Temperature (°F)	326
Gas Turbine Ouput (MW)	60.0
Steam Turbine Output (MW)	8.0
Auxiliary Requirements (MW)	7.0
Gasifier Cold Gas Efficiency	0.72
Net Electric Generated (MW)	61.0
Net Export Steam Flow (lb/hr)	274,200
Thermal-to-Electric Ratio	1.61
Overall Efficiency	0.55

NOTES:

1. Includes both steam and electric driven auxiliaries, except for gas compressor and feed pump drives.
2. Includes by-product tar and oil fired boiler at 340 psig, sat.

HEAT AND MASS BALANCE DIAGRAM
WELLMAN-GALUSHA GASIFIER

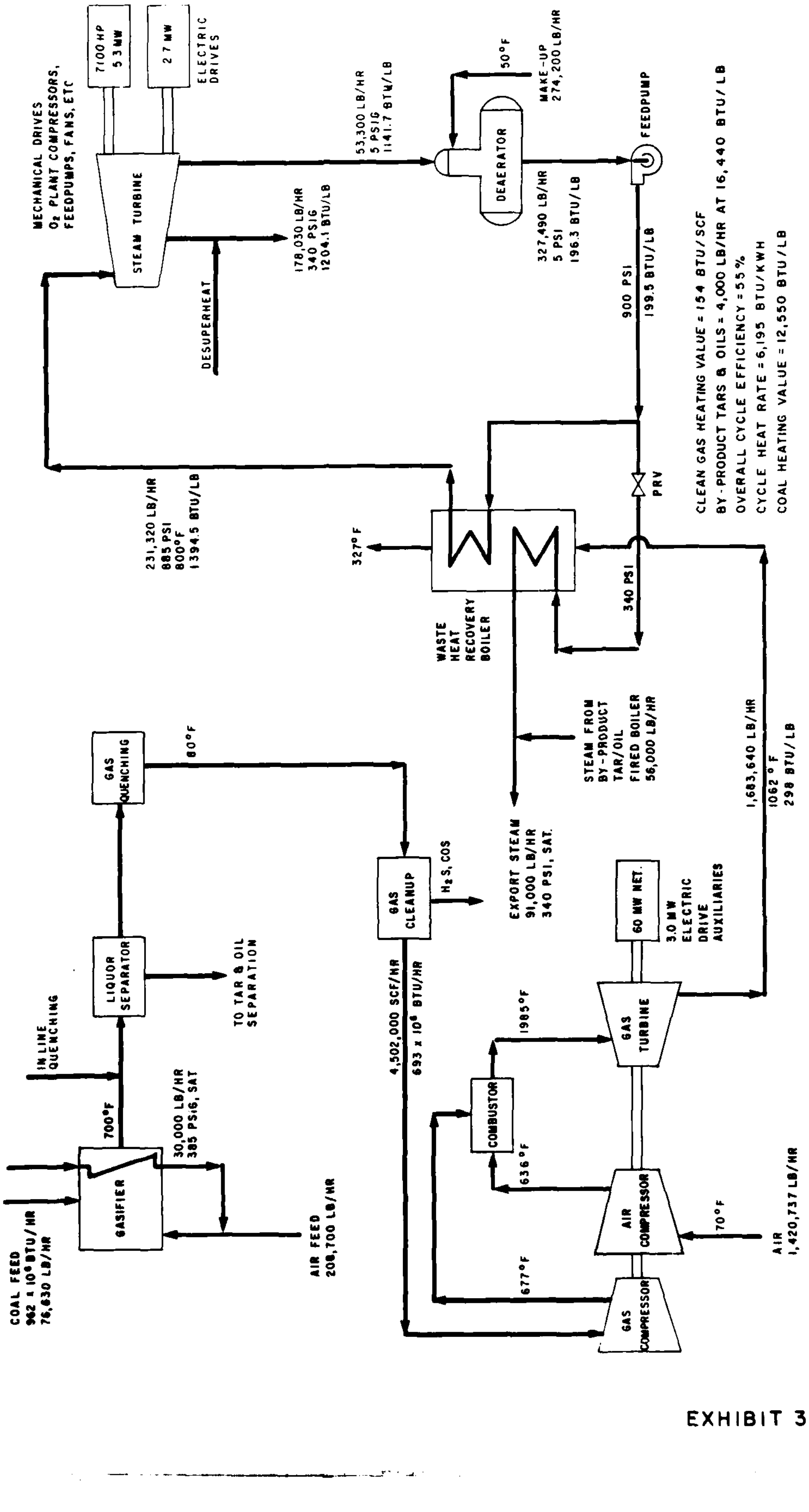


EXHIBIT 3-22

Auxiliary Firing

Auxiliary coal gas firing is applied to the Lurgi gasifier in an effort to improve its performance and increase the steam to electric ratio. Results are shown in Table 3-9. It is apparent that the significant increase in coal consumption probably does not warrant such operation.

Varying Gas Turbine Back Pressure

The effect of varying gas turbine exhaust pressure for a Koppers gasifier is shown in Table 3-10. As can be seen raising the back pressure by 5 psia increases the thermal-to-electric ratio from 1.98 to 2.74. Note the following:

- In both cases the computer program sets the gas turbine output at 60 MW. In reality, the gas turbine output will decrease with a corresponding increase in export steam flow to a new thermal to electric ratio of 2.74.

- Gas turbine performance was held constant. In reality there will be a slight degradation in performance so that the ratio of 2.74 will be met at a back pressure somewhat under 20 psia.

Gas Turbine Inlet Temperature

As pointed out in Section 3.2, the gas turbine cycle efficiency can be increased by raising the gas turbine inlet temperature. State-of-the-art turbine blades limit the gas temperature to roughly 2000°F A considerable effort is currently being undertaken to increase this limit. Goals of the program are to reach turbine inlet temperatures of 2600°F.

The effects of gas turbine inlet temperature on the combined cycle performance are shown in Table 3-11. For use in an industrial cg/cc system, the performance improvements to be obtained from these higher temperatures are not significant enough to warrant awaiting their development. Note that this would not

TABLE 3-9

LURGI GASIFIER PERFORMANCE
WITH AUXILIARY FIRING

	Base Case	Auxiliary Firing
Energy Input (10^6 Btu/hr)	899	1,092
Gasifier Oxygen Feed (lb/hr)	25,900	29,600
Gasifier Steam Feed (lb/hr)	130,000	167,000
Gasifier Jacket Steam (lb/hr)	50,000	60,000
Clean Gas Flow Rate (10^3 scfd/hr)	2,350	2,870
Clean Gas Heating Value (Btu/scfd)	282	282
By-Product Tar and Oils (lb/hr)	4,000	4,850
Throttle Steam Pressure (psig)	885	885
Throttle Steam Temperature (°F)	800	800
Throttle Steam Flow (lb/hr)	198,600	320,800
Gas Turbine Output (MW)	60	60
Steam Turbine Output (MW)	4	7.3
Auxiliary Requirements (MW)	7	8.5
Gasifier Cold Gas Efficiency	.74	.74
Net Electric Generated (MW)	57	58.8
Net Export Steam Flow (lb/hr)	221,500	348,400
Thermal-to-Electric Ratio	1.37	2.09
Overall Efficiency (%)	.51	.56

NOTES:

1. Includes both steam and electric driven auxiliaries, except for feed pump drives.
2. Includes by-products tar and oil fired boiler at 340 psig, sat.

TABLE 3-10

EFFECT OF GAS TURBINE EXHAUST PRESSURE
KOPPERS GASIFIER

	15 psia	20 psia
Energy Input (10^6 Btu/hr)	902	1,136
Gasifier Oxygen Feed (lb/hr)	47,100	59,300
Gasifier Steam Feed (lb/hr)	15,000	19,000
Gasifier Jacket Steam (lb/hr)	22,000	24,000
Clean Gas Flow Rate (10^3 scfd/hr)	2,350	2,960
Clean Gas Heating Value (Btu/scfd)	290	290
Throttle Steam Pressure (psig)	885	885
Throttle Steam Temperature (°F)	800	800
Throttle Steam Flow (lb/hr)	313,900	323,900
Stack Outlet Temperature (°F)	327	325
Gas Turbine Output (MW)	60	60
Steam Turbine Output (MW)	10.1	14.5
Auxiliary Requirements (MW)	13	16.4
Gasifier Cold Gas Efficiency	.75	.75
Net Electric Generated (MW)	57.1	58.1
Net Export Steam Flow (lb/hr)	321,100	452,000
Thermal-to-Electric Ratio	1.98	2.74
Overall Efficiency (%)	.64	.65

NOTES:

1. Includes both steam and electric driven auxiliaries, except for gas compressor and feed pump drives.

TABLE 3-11

INTEGRATED COMBINED CYCLE PERFORMANCE
EFFECT OF GAS TURBINE INLET TEMPERATURE

	1700°F	2000°F (Base)	2300°F	2600°F
Energy Input (10^6 Btu/hr)	935	902	889	886
Gasifier Oxygen Feed (lb/hr)	48,800	47,100	46,400	46,200
Gasifier Steam Feed (lb/hr)	15,600	15,000	14,800	14,600
Gasifier Jacket Steam (lb/hr)	23,000	22,000	21,000	20,000
Clean Gas Flow Rate (10^3 scfd/hr)	2,435	2,350	2,315	2,306
Clean Gas Heating Value (Btu/scfd)	290	290	290	290
Throttle Steam Pressure (psig)	885	885	885	885
Throttle Steam Temperature (°F)	800	800	800	800
Throttle Steam Flow (lb/hr)	260,000	313,900	330,000	349,900
Gas Turbine Output (MW)	60.0	60.0	60.0	60.0
Steam Turbine Output (MW)	8.8	10.1	10.5	11.3
Auxiliary Requirements (MW)	13.5	13.0	13.0	13.0
Gasifier Cold Gas Efficiency	0.75	0.75	0.75	0.75
Net Electric Generated (MW)	55.3	57.1	57.5	-
Net Export Steam Flow (lb/hr)	301,000	321,100	335,600	346,000
Thermal-to-Electric Ratio	1.92	1.98	2.06	2.09
Overall Efficiency (%)	0.58	0.64	0.67	0.69

NOTES:

1. *Includes both steam and electric driven auxiliaries, except for gas compressor and feed pump drives.*

apply for utility applications where high exhaust gas temperature and other cycle enhancements such as steam reheat would combine to increase efficiency.

3.6 Cycle Performance Summary and Conclusions

Cycle performance evaluations were carried out for the six commerically available gasifiers. Base case comparisons for a 60 MW gas turbine are shown in Table 3-12. The following conclusions can be made:

- the entrained bed gasifiers have higher overall efficiencies than the fixed and fluidized bed gasifiers.

- the fixed bed provides comparatively small steam flows for cogeneration; with the two-stage Woodall-Duckham gasifier performing better than the single-stage Lurgi. This is primarily due to the low waste heat recovery in the gas cooling stage and the high quantity of steam required as a gasifier reactant.

- the fluidized bed with its carbon carryover and char formation has a high coal use.

- the cycle configuration with the highest overall efficiency for all gasifiers includes a dual pressure boiler for maximizing waste heat recovery.

- for atmospheric gasifiers where gas compression is required, a gas driven gas compressor allows for a better match between the gas and steam turbine outputs.

- the Texaco gasifier has the highest overall efficiency, and as shown in Exhibit 3-23, its thermal-to-electric ratio best matches the base requirements.

TABLE 3-12

INTEGRATED COMBINED CYCLE
PERFORMANCE SUMMARY

	Lurgi	Woodall-Duckham	Wellman-Galusha[3]	Winkler	Koppers-Totzek	Texaco
Energy Input (10^6 Btu/hr)	899	958	962	1,070	902	865
Annual Coal Use (tons)[1]	320,400	341,400	342,800	381,300	321,500	308,300
Net Electric Generated (MW)	57.0	56.9	61.0	58.4	57.1	56.2
Net Export Steam Flow (lb/hr)[2]	221,500	274,600	274,200	325,300	321,100	330,100
Thermal-to-Electric Ratio	1.37	1.70	1.61	1.97	1.98	2.07
Overall Efficiency (%)	51	54	55	55	64	68
Heat Rate (Btu/kWh)	6,690	6,320	6,210	6,210	5,330	5,020

NOTES:

1. Standard coal at 12,290 Btu/lb; 100% availablilty.
2. Includes by-products, tar and oil, fired in a boiler at 340 psig, saturated.
3. Air-blown gasifier.

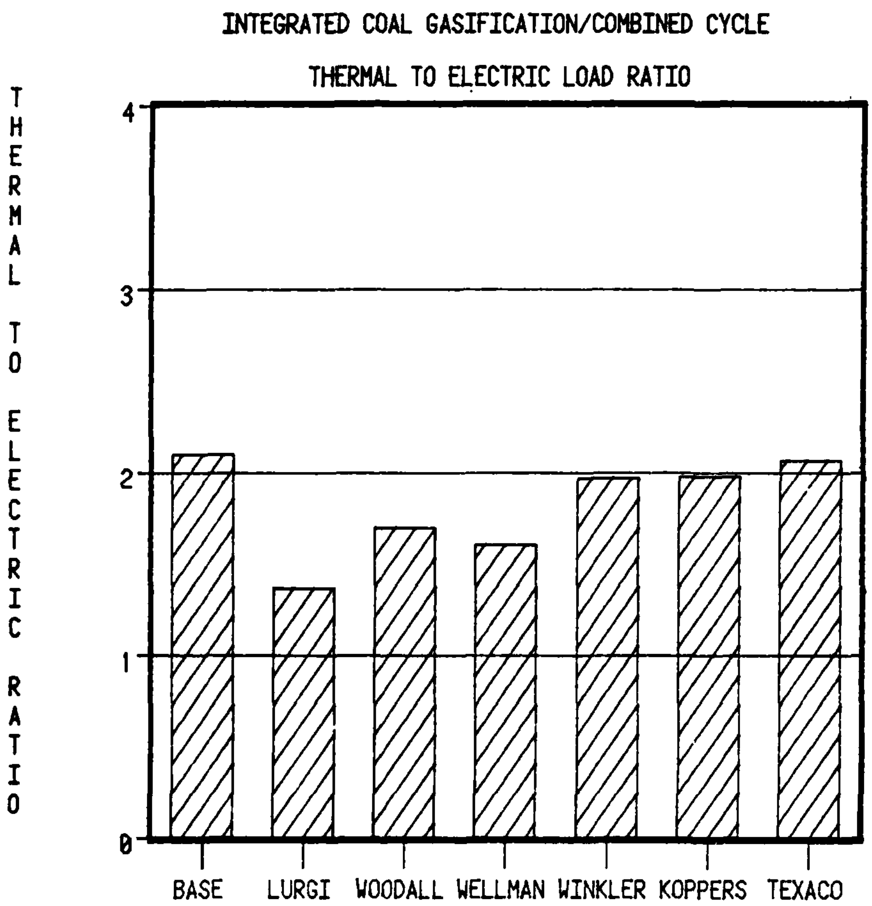

INTEGRATED COAL GASIFICATION/COMBINED CYCLE
THERMAL TO ELECTRIC LOAD RATIO
THERMAL TO ELECTRIC RATIO
4
3
2
1
0
BASE
LURGI
WOODALL
WELLMAN
WINKLER
KOPPERS
TEXACO

EXHIBIT 3-23

In addition, a significant number of cycle improvements were assessed. These included:

- improvements to fixed bed gasifier performance by providing auxiliary coal gas firing; results indicate that the significant increase in coal consumption probably does not warrant such operation.

- effects of combustion inlet temperature for the gas turbines; the performance improvements at temperatures up to 2600°F are not significant for industrial-based cg/cc systems.

- variations in gas turbine back pressure, i.e., under-expanding to maintain high exhaust temperatures and thereby increasing steam output; such back pressure control can provide a significant means for varying system thermal-to-electric ratio.

For generic performance and sizing purposes, we have developed unit output curves for the gas turbine, steam turbine and net electric output, steam export, overall efficiency, and thermal-to-electric ratio. These results, all on a unit ton of coal basis, are shown in Exhibits 3-24 through 3-28, using a typical atmospheric entrained bed gasifier. From this we may observe the significant effects of back pressure control.

EFFECT OF COMBUSTION TEMPERATURE AND GAS
TURBINE BACKPRESSURE ON TURBINE OUTPUT

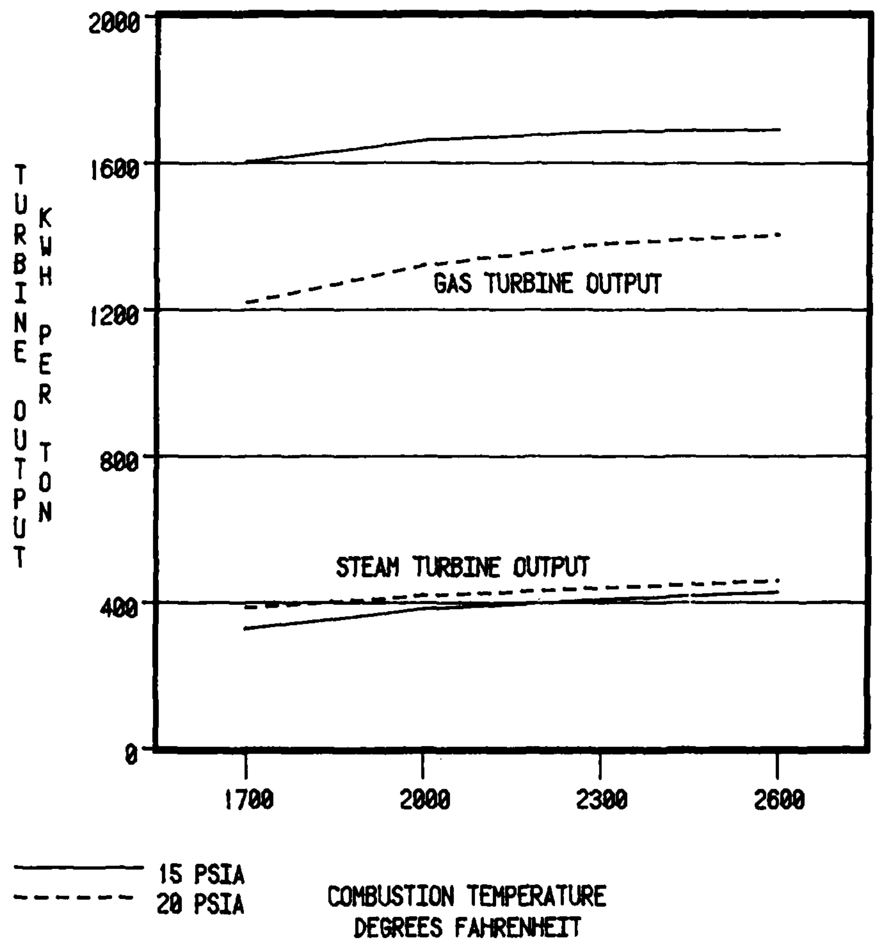


EXHIBIT 3-24

EFFECT OF COMBUSTION TEMPERATURE AND GAS TURBINE BACKPRESSURE ON NET ELECTRIC OUTPUT

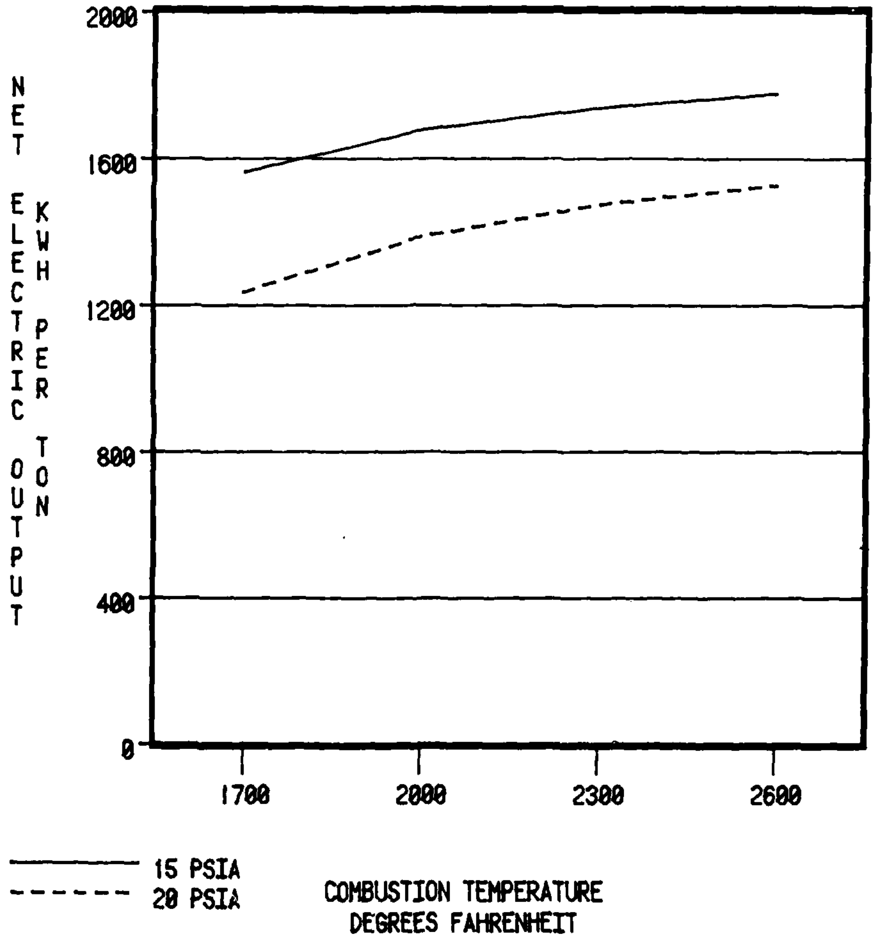


EXHIBIT 3-25

EFFECT OF COMBUSTION TEMPERATURE AND GAS TURBINE BACKPRESSURE ON STEAM EXPORT

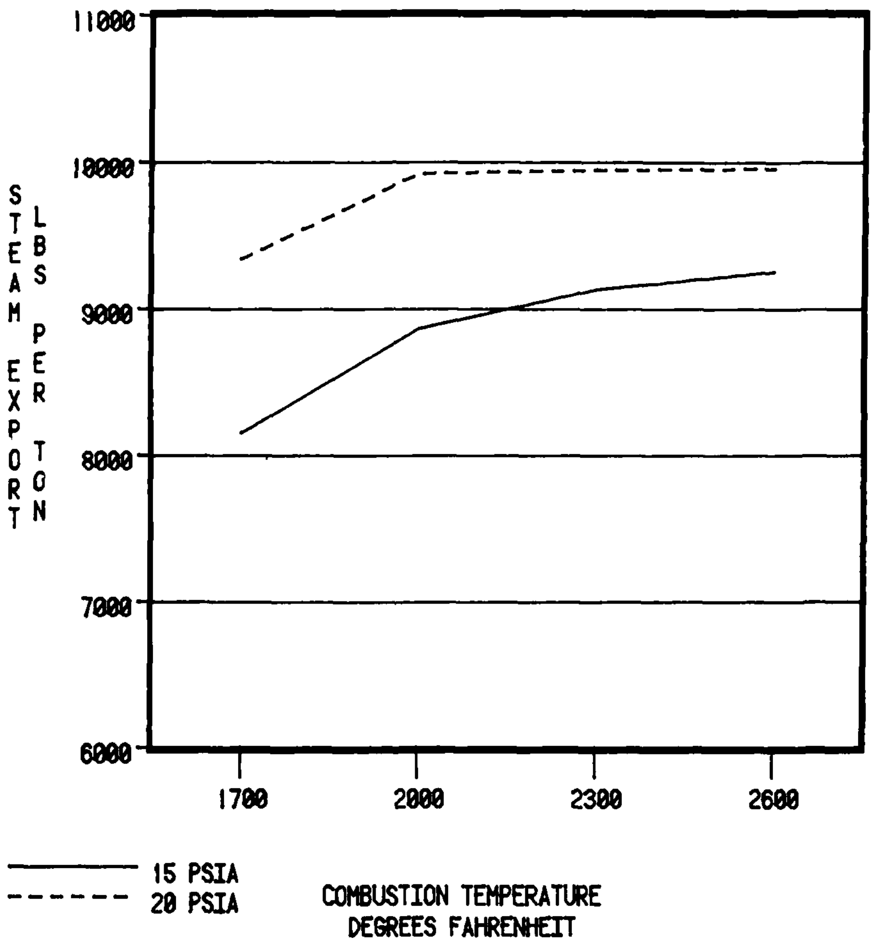


EXHIBIT 3-26

EFFECT OF COMBUSTION TEMPERATURE AND GAS

TURBINE BACKPRESSURE ON OVERALL EFFICIENCY

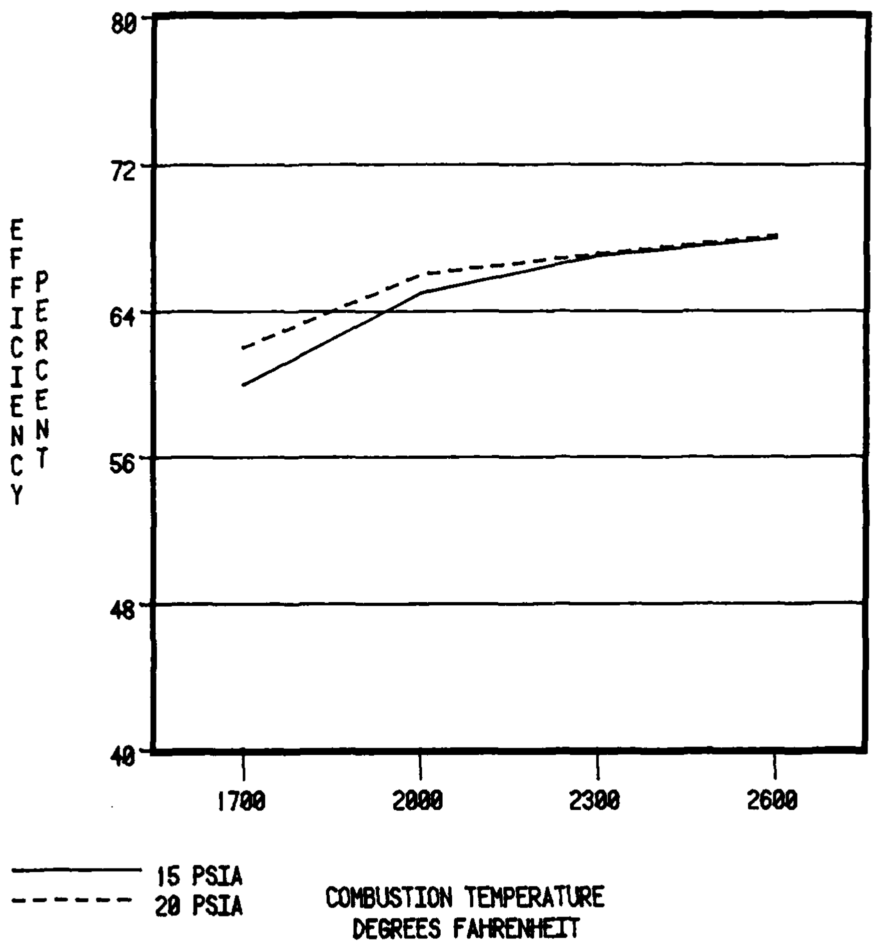


EXHIBIT 3-27

EFFECT OF COMBUSTION TEMPERATURE AND GAS TURBINE BACKPRESSURE ON THERMAL TO ELECTRIC RATIO

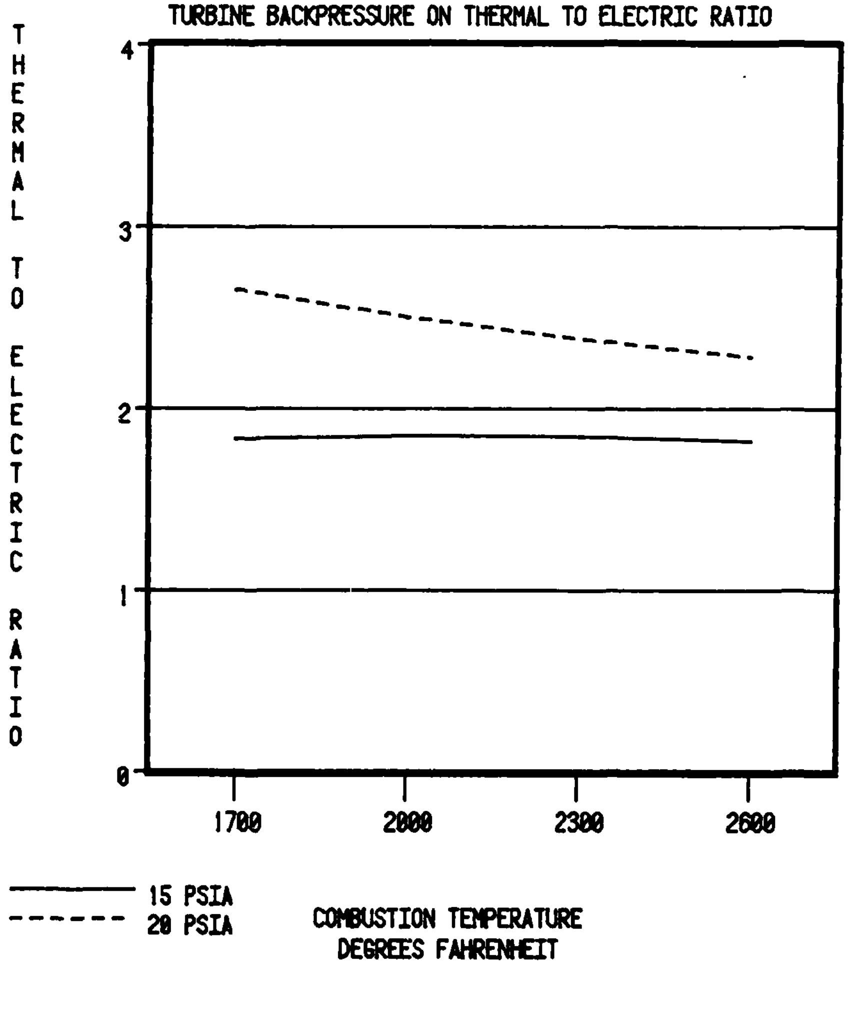


EXHIBIT 3-28

3.7 REFERENCES

3-1 Hilt, M.B. and R.A. Farrell, "Low Btu Gas Capabilities", Gas Turbine Reference Library, GER-3092, General Electric Company, Schenectady, New York, 1979.

3.2 "Engineering News and Trends", Gas Turbine World, November, 1980.

3.3 Foster, R.W., "The Integration of Gasification With Combined Cycle Power Plants", Combustion, December, 1979.

3-4 "SYNTHA II Power Plant Design and Surveillance", Application Reference Manual Control Data Corporation, Minneapolis, Minnesota, 1978.

3-5 Hartman, H.F., "Low Btu Coal Gasification Processes", Oak Ridge National Laboratory, Oak Ridge, Tennessee, 1978.

4.0 ENVIRONMENTAL CONTROLS FOR THE COMBINED CYCLE

With the nature and performance of the gasification/combined cycle system complete, it is now pertinent to turn our attention to an analysis of the sources of effluents for the process and to a presentation of the associated cleanup technologies. To set the stage for this discussion, reference should be made to Exhibit 4-1 which, drawing on the flow diagram of Section 3, shows the various effluent streams.

As may be observed in the exhibit, and for convenience in the presentation that follows, we treat the cleanup sequence in four basic steps:

- coal storage, treatment and processing controls
- gasification controls
- gas purification controls
- gas turbine combustor and exhaust stack controls

Furthermore, we will emphasize those elements which are unique to the gasification/combined cycle process accepting as standard traditional coal cleanup technologies. In doing so we maintain the guideline of only looking at commercially available equipment. Selection criteria are the same as we used to screen the available gasifiers, viz:

- status
- technical factors
- capacity
- data availablility

At the end of this section, we also provide a discussion of the matching of the gasifiers selected in Section 2.0 to the control processes described here. We should recognize, however, that gasifier manufacturers generally recommend specific techniques to match their system requirements; thus our presentation should be considered representative of the

SOURCE OF AIR, WATER AND SOLID EFFLUENTS

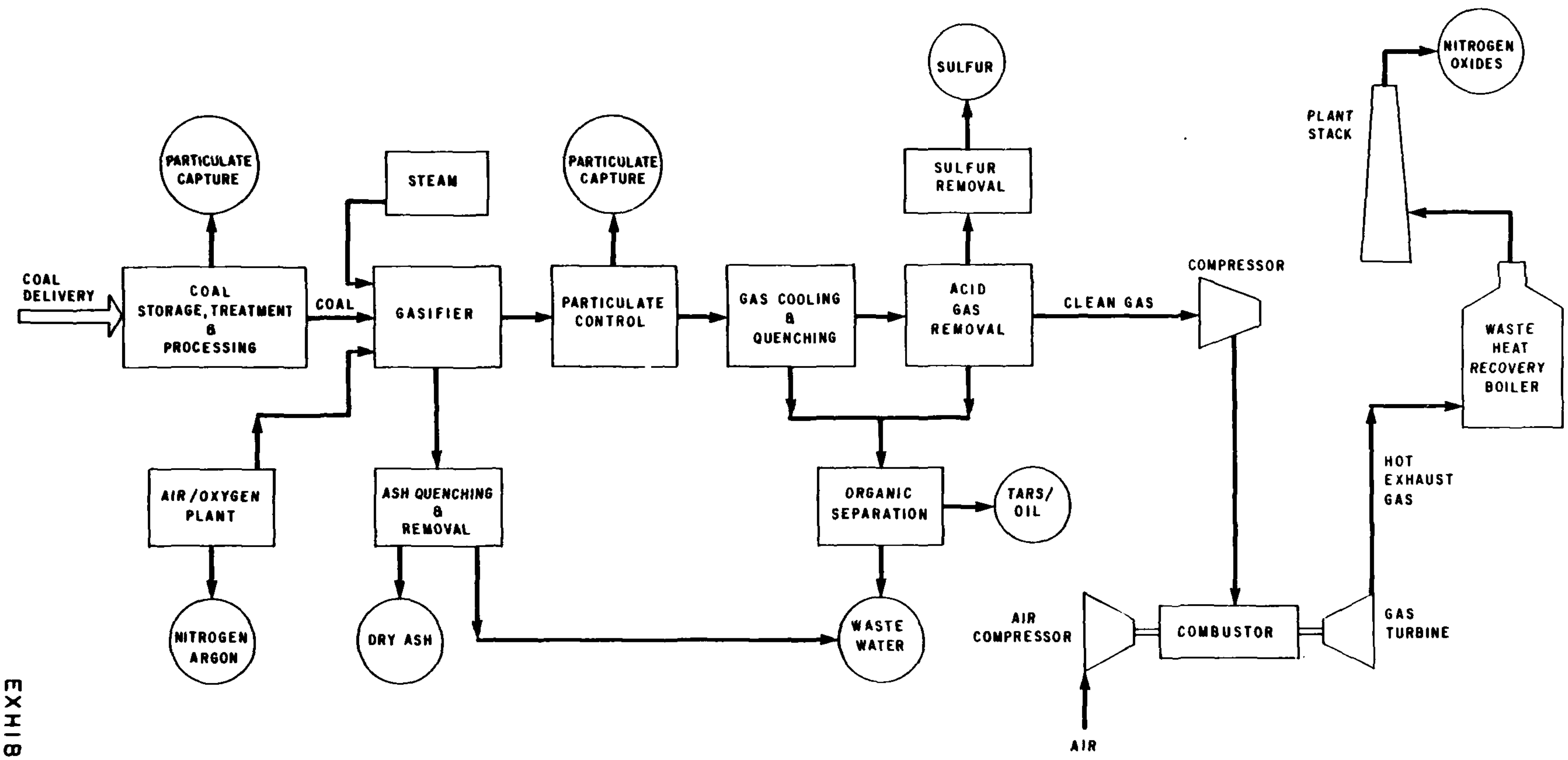


EXHIBIT 4-1

considerations involved. Detailed matching awaits the formal submission of budget estimates by the gasifier companies who typically provide coal-pile-to-clean-gas systems.

In order to provide the necessary background to assess the control techniques, some discussion of environmental standards is required. We do this first in Section 4.1; this is followed by the details of the control systems in Sections 4.2 - 4.5.

4.1 Environmental Standards

Given the complexity of the basic system under consideration and its relative newness with regard to operations, at this time it is not clear which of the many regulations a gasification/combined cycle system might have to meet. For the purpose of this discussion, then, we review what appears applicable, citing in as much detail as necessary the substance of the laws, to determine their relevance.

Consider first the Federal air pollution regulations. In this regard the U.S. Environmental Protection Agency (USEPA) has promulgated several New Source Performance Standards (NSPS) which should be addressed:

Subpart D - "Standard of Performance for Fossil-Fuel-Fired Steam Generators for Which Construction is Commenced After August 17, 1971." From the definitions in Subpart D, the affected facilities are basically fossil-fuel-fired steam generators of greater than 250 million Btu per hour heat input. This subpart does not address gas turbine or coal gasification equipment. Further, since the limited supplementary firing considered for the cycle in this study is substantially below the heat input limits, the subpart is not applicable to the coal gasification/combined cycle.

Subpart De - "Standards of Performance for Electric Utility Steam Generating Units for Which Construction is Commenced After September 18, 1978." This subpart applies to electric utility combined cycle gas turbines that are capable of combusting more than 250 million Btu per hour heat input of fossil fuel in the steam generator. Only emissions resulting from combustion of fuels in the steam generating unit are subject to this subpart. Again, since the supplementary firing is small relative to the standard, Subpart De also does not apply.

Even though Subparts D and De, by their definitions of system inclusion, do not apply to the gasification/combined cycle here, it is perhaps worthwhile to summarize the standards to be derived from these regulations. We do so (1) to offer some idea of the amount of control generally required of the size of facilities at Sewells Point, (2) to provide required input for other cycles that might be of interest in this study (see Section 5.0) and (3) to indicate that there has been some discussion that these might ultimately become the standards for gasification systems. Finally, in reviewing these regulations, we should keep in mind that the several turbine manufacturers generally specify NSPS, Subparts D and De, as the limit of acceptable input to their combustors. For the record then, the levels of cleanup required for Subparts D and De are shown in Table 4-1.

Subpart GG - "Standards of Performance for Stationary Gas Turbines." This subpart is applicable to all gas turbines with heat input at peak load equal to or greater than 10.15 million Btu per hour. It is this subpart then which applies to the combined cycle. In Subpart GG standards for nitrogen oxides and sulfur dioxide are promulgated. These we show in Table 4-2. With regard to the former, while this may appear

TABLE 4-1

NEW SOURCE PERFORMANCE STANDARDS

Subpart D - Fossil-Fuel Fired Steam Generators
Subpart De - Electric Utility Steam Generating Units

Performance Standards for Coal and Coal Derived Fuels

Effluent	Subpart D	Subpart De
Particulate Matter	0.10 $lb/10^6$ Btu heat input	99% reduction of uncontrolled emission with maximum of 0.03 $lb/10^6$ Btu heat input
Opacity	No greater than 20% opacity	No greater than 20% opacity
Sulfur Dioxide	1.2 $lb/10^6$ Btu heat input	90% reduction of uncontrolled emission with maximum of 1.2 $lb/10^6$ Btu heat input, or 70% reduction when emissions are less than 0.6 $lb/10^6$ Btu heat input
Nitrogen Oxides	0.70 $lb/10^6$ Btu heat input	25% reduction of uncontrolled emissions with maximum of 0.50 $lb/10^6$ Btu heat input

TABLE 4-2

NEW SOURCE PERFORMANCE STANDARDS

Subpart GG - Stationary Gas Turbines

Effluent	Performance Standard
Nitrogen Oxides	75 ppm
Sulfur Dioxide	150 ppm or combustion of any fuel which does not contain sulfur in excess of 0.8 percent by weight

to be a severe standard, current turbine technology indeed permits satisfaction of this regulation (see Section 4.5). Note that the sulfur dioxide standard is given two ways: input or output. Most cleanup technologies permit meeting the input side easily and thus sulfur control is not a problem here (see Section 4.4).

In addition to meeting the Federal regulations, every proposed system must also satisfy State of Virginia Air Pollution Control Regulations. For purposes of the various parts of the state codes, Sewells Point is located in Region 6 - Hampton Roads Intrastate Air Quality Control Region of Virginia. As might be expected, and similar to the Federal laws, no specific mention of gasification/combined cycle plants may be found in the Virginia laws. However, as before, we cite the relevant fossil fuel regulations and assume their applicability here.

The State standard for particulates indicates that for facilities with total capacity between 10 million and 10 billion Btu per hour, the maximum allowable emission rate, E, in pounds of particulate per million Btu input, shall be determined by the following equation:

$$E = 1.0906\ H^{-0.2594}$$

where H is total capacity in millions of Btu per hour. The standard for sulfur dioxide limits emission, S, in lbs/hr to

$$S = 2.64\ K$$

where K is the actual heat input at total capacity expressed in Btu x 10^6 per hour. Taking the gas turbine as the combusting source, indicates that $H \approx 700$ or $K \approx 700$ and that

$$E = 0.20 \text{ lbs of particle}/10^6 \text{ Btu}$$

and

$$S = 1850 \text{ lbs/hr}$$

Clearly, the Federal regulations are more stringent and thus they would apply.

Since the pollution streams, see Exhibit 4-1, contain solid and liquid wastes, it is also important to address the regulations for those effluents. In this case, as we will see, in the sections to follow, those waste schemes are not a problem and standard disposal techniques, e.g., landfill, may be applicable.

4.2 Coal Storage, Treatment and Processing Controls

With the regulations established we now turn our attention to the various effluent streams of the gasification/combined cycle system shown in Exhibit 4-1. The first phase of the overall process is in fact common to all coal-firing systems. Coal is delivered to the site, unloaded, conveyed to storage, reclaimed from storage, processed (sized) and then delivered to the system. Effluents here arise from both the handling and storage. In the former, we must deal mainly with the problem of fugitive dust; in the latter, dust as well as coal pile runoff or leachate control is required.

Techniques for handling these emissions are now standard in the industry and include:

- Baghouses for unloading buildings.
- Spraying of the coal prior to storage for dust suppression.
- Lining of open storage areas to provide for systematic leachate capture.
- Covered conveyor systems for dust supression.
- Collection hoods and ducts for the sizing process.

References to the standard literature in this regard are sufficient for the purposes of this study, see e.g., References 1, 2 and 3.

4.3 Gasification Control

As shown in Exhibit 4-1, coal, steam and air or oxygen for either low or medium Btu gas enters the gasifier. The basic output is the stream of raw gas with effluent streams arising from the delivery of coal, the generation of steam, the generation and delivery of the oxidant and the removal of ash from the gasifier. As a general statement, we may note that control technologies for this step do not differ widely from those controls for traditional coal-fired boiler systems.

The nature of the outputs depends heavily on the type of gasifier selected as may be seen from a review of the data in Section 2 or Appendix A. Some of the more important parameters affecting the effluent are:

Coal Feed Stock - First, increases in the amount of volatile matter tend to increase the amounts of methane, tars and oils in the raw gas. Further, coal feeding techniques also affect the quantity of tars and oils. Gasifiers which feed coal at the top of the bed (fixed bed types) tend to produce larger amounts of tars and oils than do the fluidizing types. Entrained beds, which use pulverized coal, produce little or no tars and oils.

Pressure - Increasing pressure tends to favor production of higher heating value raw gases with increased generation of methane and carbon monoxide.

Temperature - Raw gas composition is affected by gasification temperature especially in fixed and fluidized bed types. As the temperature increases thermal cracking of tars, oils, phenols and hydrocarbon increases. Ash handling problems likewise increase with a rise in temperature.

Now consider in detail the various elements in the gasification process and their effluent control (see References 4 and 5). We first treat the coal handling system which is a major source of potential air emissions. They may contain raw gasifier product gas components, coal or ash dust, and, in pressurized systems, pressurizing gas components.

Some coal dust will always be generated as a result of transporting coal to the gasifier feed hopper. In most systems, the use of a covered coal transportation system along with gas collection ducts and particulate removal equipment would be acceptable to limit these emissions.

There are four general types of coal feeding devices which are in widespread use:

- Lock Hoppers
- Rotary Feeders
- Screw Feeders
- Slurry or Entrained-Flow Injection Devices.

Lock hoppers and slurry injection devices are used to feed coal to high-pressure gasifiers while lock hoppers, rotary feeders and screw feeders are used to feed coal to atmospheric pressure gasifiers.

Vent gases from lock hoppers and rotary feeders used on atmospheric pressure gasifiers will contain raw gasifier product gas components. A suitable purge or blanketing gas into the gasifier provides control here. The composition of air emissions from lock hoppers used on pressurized gasifiers will depend on the method of pressurizing the lock hopper; for example:

- Prior to dumping the coal from the lock into the gasifier, the lock may be pressurized.

- If the pressurizing gas is added continuously, the gas remaining in the lock will have approximately the same composition as the pressurizing gas.

- If no gas is added as the coal is dumped, raw gas from the gasifier will fill the void space created as the coal falls into the gasifier.

- If no pressurizing gas is used, the lock will fill with raw gas as the coal is dumped into the gasifier, and the gas remaining in the lock will be composed of raw gas.

The Texaco gasifier uses a liquid slurry for feed of their pulverized coal. With the use of a liquid slurry, there is usually an efficiency penalty which results from the vaporization of the coal carrier liquid.

All the gasifiers under consideration require steam as part of their input streams. In the combined cycle application discussed in this report, we envision that the steam required for use in the gasifier will either be developed by the high temperature of the gasifier itself or developed by the waste heat recovery from the gas turbine exhaust. Therefore, no special environmental requirements arise.

The introduction of the oxidant gives rise to only minimal problems. For production of low Btu gas, an air stream is required for which there would be no effluents. Current technology for generating the high purity oxygen stream

needed to produce medium-Btu gas involves the use of cryogenics. In such a unit, inlet air is compressed, pre-cooled and liquefied by flash cooling and by contact with cold product gas streams. The only potential direct emissions from this processing step are the high purity nitrogen and argon streams which are produced as by-products of the liquid air fractionation step. For the system here, it is likely that these streams would be vented to the atmosphere.

Finally in the gasification process, we consider the ash system. The initial requirement is the removal of hot ash or slag from the gasifier and the cooling or quenching of that material, usually with water. We discuss here the problems of fixed- and fluid-beds. In entrained-bed systems the ash must be separated from the product gas; this is addressed in connection with the gas purification requirements.

The ash handling devices used by fixed- and fluid-bed gasifiers for this dry ash include:

- Water-Sealed Ash Pans.
- Screw Conveyors.
- Lock Hoppers

Quench systems are used to cool the ash or slag removed directly from the gasifier. The quench system includes a pressure let-down device when it is used with a high pressure gasifier.

Air emissions from water sealed ash pans and other quench systems will contain volatile materials that evaporate from the ash pan water. These volatiles may either be components which enter the system with the ash pan makeup water or they may be products of reactions between the ash pan water and the hot gasifier ash. The composition of the gasifier ash will obviously have a significant affect upon the quantities and compositions of the volatile materials released by this

mechanism. Very little volatile material should be derived from the quenched ash leaving a fixed-bed gasifier. There is a greater potential for the release of hydrocarbons from the ash leaving a fluid-bed gasifier because this material is more "char-like" than the more completely oxidized residue of a fixed-bed gasification process.

The composition of emissions from lock hoppers will be dependent on its mode of operation. For atmospheric pressure gasifiers which discharge a dry, unquenched ash, the emissions will consist of steam and air (or oxygen), and ash particles. If the ash is quenched prior to discharge from the lock hopper, products of reactions between the quench water and the hot gasifier ash may be present in the air emissions. Control technologies that are applicable to the control of air emissions from ash handling systems are similar to those employed to control coal feeding system emissions and are similar to traditional coal firing technologies. Containment and collection of particulate-laden air followed by processing in a suitable particulate control process will be needed with dry ash systems where ash dust emissions are a problem. This control usually involves the use of quench or sluicing system makeup water that does not contain hazardous materials that are or will form volatile components upon contact with hot gasifier ash.

An ash quenching and/or sluicing system, if used, is a major source of potential liquid effluents from the coal gasification operation. Ash removal devices which discharge a dry, unquenched ash do not produce liquid effluents. The liquid effluents produced by ash quenching sluicing systems will contain varying amounts of suspended ash or slag particles, and soluble components leached from the ash as well as components initially present in the quench water makeup.

All ash removal devices are sources of solid wastes since the mineral matter in the gasifier ash or slag is a solid waste. In addition to the mineral matter from the feed coal, coal feed additives and unreacted coal may also be present in this solid waste stream. Components present in the quench water input may also be present in the ash or slag. The ultimate composition of the waste ash or slag will depend upon the gasifier type, its operating conditions, the coal feed-stock and additive compositions, and the makeup quench water composition.

4.4 Gas Purification Controls

The purpose of the gas purification operation is to remove constitutents such as particulates, tars, oils and acid gases from the raw product gas and generate the clean gas for use in the turbine combustor. It is this step which is unique to the gasification process and for which we direct considerable attention. There are basically three steps in the sequence necessary to produce this clean gas:

- Particulate removal.
- Gas quenching and cooling.
- Acid gas removal and sulfur capture.

As we treat these steps in turn, we will see that not all are pertinent to every gasifier (see References 4-8).

Particulate Removal

Removal of coal dust, ash and tar aerosols entrained in the raw product gas leaving the gasifier is the primary function of this step. Specific processes commonly used to accomplish this are:

- Mechanical Collectors
- Electrostatic Precipitators (ESP), and
- Scrubbers

We may observe that these are not different from traditional particulate control technologies.

Mechanical collectors remove particulate matter from gas streams by the actions of physical forces such as gravity, centrifugal force, impingement and diffusion. Three types of mechanical collectors which are widely used to control particulate emissions from industrial processes include:

- Settling Chambers
- Cyclones
- Baghouses

The effectiveness of each of these types of collectors depends mainly upon the size distribution of the particulate matter and the flow rate and physical properties of the gas stream. Filters generally provide better collection efficiencies than the other two types of collectors, especially if very small particles (<5 μm) must be collected. However, cyclones are used generally as the initial cleanup step on most operating commercial gasifiers.

Electrostatic precipitators (ESP's) remove particulate matter from gas streams by the action of an electrical field on charged particles. Two types of ESP's (high- and low-voltage) are commercially available. However, the low voltage types were not designed with coal fired systems as their base; they do not, therefore, play a role here.

High-voltage ESP's are used when predominantly small particles (<20 μm) must be removed from large volumes of gas. Collection of particulate matter by high-voltage ESP's involves three basic steps:

- Transmitting an electrical charge to the particulate matter.

- Collecting the charged particles on a grounded surface.
- Removing the collected particulates from the precipitator.

Because of high collection efficiencies associated with high-voltage ESP's, they are generally applicable to control of particulate emissions from coal gasification plants.

Scrubbers use a liquid, usually water, either to remove particulate matter from a gas stream by direct contact or to increase collection efficiency by preventing re-entrainment of the collected particles. There are many types of wet collectors, all of which are some variation of a srpay chamber or a wet scrubber. Wet scrubbers are relatively high energy using devices. For this reason, wet scrubbers often do not compare favorably with mechanical collectors or ESP's, in applications where particulate removal is the only control required as is the case here. This is in distinction to their use as stack cleaning devices for coal fired boilers. Therefore, while a baghouse or an ESP might be better suited to the removal of coal and ash dust from gas streams which are collected in the vicinity of solids handling operations, wet scrubbers appear to have application in the removal of particulates and SO_2 from on-site combustion stack gases.

The various particulate control measures are compared in Table 4-3. For the input coal and the various gasifiers under consideration here, cyclones followed by either baghouses or ESP's are most appropriate.

Gas Quenching and Cooling

In gas quenching and cooling, tars and oils are condensed and particulates and other impurities such as ammonia are scrubbed from the raw product gas. Quenching involves the

TABLE 4-3

PARTICULATE CONTROL DEVICES

DEVICE	ADVANTAGES	DISADVANTAGES	COMMENTS
MECHANICAL COLLECTORS			
1. Settling Chambers	• Low Energy Devices	• Large size due to high residence time and low flow requirements • Low removal efficiency for fine particulates	• Does not appear to be well suited to coal gasification plant particulate control applications.
2. Cyclones	• Mechanically Simple • Low Cost	• Not an effective collector of fine particulates	• Is a low energy device for large particulates, but requires higher energy dissipation to remove fine particulates.
3. Baghouses	• High collection efficiency	• Caking/Plugging problems incurred with wet, saturated gases	• Medium Energy Device. • Of the mechanical collectors, probably the best suited to the control of gasification plant coal and ash dust emissions.
ELECTROSTATIC PRECIPITATORS			
1. High-Voltage	• High collection efficiency • Suitable for fine particulate collection • High gas flows can be treated • Can collect liquid and solid particulate matter	• High voltages required • Sticky liquids can collect on the collection electrode and decrease efficiency	• Very effective device for removing fine particulates from large gas flows. Typical applications have been on coal fired boiler flue gases.
2. Low-Voltage	• Low voltage required	• Cannot handle solid or sticky liquid particulate matter	• Since only application is to non-sticky liquid particulates, this device does not appear to be suited to coal gasification plant particulate control applications.
SCRUBBERS	• High efficiency can be obtained with certain scrubber types	• Liquid wastes are produced • To obtain high collection efficiencies requires high energy dissipation	• The need for treating the resultant liquid waste detracts from wet scrubbers as a particulate-only control device.

direct contact of the hot raw gas with an aqueous or an organic quench liquor. Extensive cooling of the gas stream occurs initially, primarily through vaporization of the quenching medium. Further gas cooling can be accomplished using waste heat boilers followed by air- and/or water-cooled heat exchangers.

The choice of gas quenching and cooling processes to be used depends upon the nature of the hot raw gas and whether or not an acid gas removal process will be needed. Waste heat recovery is important and necessary (see Section 3) but limits to such recovery arise because of fouling problems due to tar and oil condensation in the waste heat boiler. The amount of cooling required is dictated by the acid gas removal process temperature constraints.

Gas quenching and cooling is a source of liquid effluents and solid wastes. The liquid effluents consist of the quench liquor and the tars and oils condensed in the quenching process. The composition and amounts of these tars and oils depends on gasifier process considerations (coal type, pressure, temperature, etc.) and the nature of the quenching medium (i.e., water or light oil). The amount of condensate produced is directly affected by the temperature to which the gas is cooled. Furthermore, this discussion is largely relevant only to fixed bed gasifier; fluidized and entrained beds do not generate significant quantities of tars and oils.

There are many steps involved in the quenching and cooling operations to develop both a suitable gas for the next step (acid gas removal) and an environmentally acceptable waste water stream. For the system under evaluation for Sewells Point, probably only the primary processes (see below) are necessary. The tars and oils captured from that stage can

probably be combusted or sold as a low grade fuel requiring additional cleanup in the resulting boiler stack. For completeness here we provide a representative discussion of all the elements in the cleanup procedure:

(1) Oil/water separators and suspended solids removal systems are the primary processes with typical techologies:

- Flocculation/Flotation: Flocculation involves the addition of chemical additives to coagulate five solids. Floatation uses air bubbles to raise the oil droplets to the water surfaces where they may be skimmed off.

- Oil/Water Separators: Gravity separators are used to remove non-emulsified oils and suspended solids.

- Filtration: Filtration reduces concentration of suspended contaminants in aqueous streams.

(2) Dissolved organic removal systems are secondary and tertiary processes with typical technologies:

- Extraction: This involves the removal of phenols by liquid-liquid chemical extraction techniques.

- Adsorption: This process is useful to recover phenols and generally removes dissolved organics.

- Biological Oxidation: Bacteria and other microbes are used for removal of organics; best suited as a tertiary treatment scheme.

(3) Dissolved inorganics removal systems are also secondary and tertiary processes with typical technologies:

- Acid Gas Stripping: The waste stream is contacted by a countercurrent of an inert gas to remove acid gases.

(4) Residual containment removal systems are the final steps and include the following typical technologies:

- Forced Evaporation: This process removes dissolved salts in the form of a concentrated sludge.

Considerable detail is available in the current literature concerning these systems (see References 4 and 5). For our purposes here, it is sufficient to summarize process information recognizing that all such systems are commercially available, all have demonstrated the control effectiveness needed to meet environmental standards and acid gas removal input conditions and all have shown good operating reliability. This summary is provided in Table 4-4. Details of representative processes are provided in Appendix C1.

There remains an analysis of the use or disposal of the collected tars and oils for the fixed bed gasifiers. Generally, the content of such tars and oils is the unburned carbon from the gasifier, particulates not captured by these controls and the trace elements originally present in the coal. Note that the energy content here is significant and can amount to nearly 10% of the coal input. For our purposes we have assumed (see Section 3) that these effluent fuels can be burned in an auxiliary but separate boiler with its own stack controls to satisfy air pollution regulations. Note that such regulations only refer to particulates and not to the trace elements generally present in the tars and oils.

TABLE 4-4

SUMMARY OF GAS QUENCHING AND COOLING CONTROL PROCESSES

TREATMENT FUNCTION	SUSPENDED SOLIDS AND OILS REMOVAL			DISSOLVED ORGANICS REMOVAL			DISSOLVED INORGANICS REMOVAL	RESIDUAL CONTAMINANT REMOVAL
	Flocculation Flotation	Oil-water Separation	Filtration	Liquid-Liquid extraction (Phenosolvan)	Activated carbon absorption	Biological oxidation (activated sludge)	Acid gas Stripping	Forced Evaporation
Coal Gas Applicability								
• Presently used	yes	yes	yes	yes	no	yes	yes	no
• Potential future use	yes	yes	yes	yes	yes	yes	yes	yes
Control Effectiveness								
• Suspended solids removal	~75%	~90%	70-75%		~90%	70%		
• Free oil removal	~97%	~90%	52-83%		~93%	80%		
• Phenol removal		~25%		>94%	99+%	95-99%	20-40%	~99+%
• Total organics removal				~90%	~90-95%	~90-95%	~90-95%	
• BOD removal	~80%	~80%	36%			~90%		
• Sulfide removal						~97%	~99%	
• NH_3 removal						15%	~90%	
• Cyanate removal					~1%	~70%		
• COD removal	80%	~50 ppm ≤	25-44%		~90%	~99.9%		
• Trace element removal	✓					✓		✓
• Total dissolved solids removal								99%
Utility Requirements								
• Steam				✓				
• Electricity	✓	✓	✓	✓	✓		✓	✓
• Cooling/backwash H_2O			✓	✓	✓	✓	✓	✓
• Fuel gas					✓	✓		
Raw Materials Required								
• Solvent				✓				
• Chemical additives	✓					✓		
Allows By-Product to be Recovered	✓	✓	✓	✓			✓	
Generates Effluents Requiring Further Control								
• Gaseous					✓		✓	✓
• Aqueous	✓			✓	✓			
• Treated effluent	✓	✓		✓			✓	
• Solid/semisolid	✓	✓				✓		✓
Process Limitation/ Sensitivity								
• Pressure level	atmospheric	atmospheric	wide range	atmospheric	atmospheric	atmospheric	atmospheric	vacuum to atmospheric
• Temperature change	✓			solvent dependent		✓	✓	✓
• pH level	✓				✓	✓	✓	
• Contaminant size distribution	✓	✓						
• Requires regeneration			✓	✓	✓			
• Adversely affected by trace elements						✓		
• Nutrients required						✓		
• Chemical additives required	✓					✓		
• Hydraulic loading	✓					✓		
Advantages	inexpensive effective	simple effective high reliability low energy use	effective maintainability inexpensive	effective	effective for low concentration pollutants wide applicability	effective for BOD & Phenols inexpensive	inexpensive proven effective	environmentally acceptable low energy use
Disadvantages	high sensitivities auxiliary requirements	large space requirements effluents require control only good for non-emulsified solids	regeneration required effluents require control	require solvent reclaim for economy high first cost	requires recovery of adsorbent for economy	high sensitivity	steam sensitive corrosive outputs	high capital investment inefficient further treatment of effluents

The economic effect here is assssed later in our comparison of fixed beds with other types.

Acid Gas Removal and Sulfur Capture

Acid gases such as H_2S, COS, CS_2, mercaptans and CO_2 are removed from the raw product gas in this step. Processes used for acid gas removal may remove both sulfur compounds and CO_2 or they may be operated selectively to remove only the sulfur compounds in cases where carbon dioxide removal is not required to meet clean gas specifications.

The processes used for acid gas removal may be divided into two general categories:

- High-temperature processes requiring minimal cooling of the feed gas before treatment; and
- Low-temperature processes requiring extensive cooling of the feed gas before treatment.

Presently there are no commercially available high-temperature processes. Therefore, our focus will be on the low-temperature systems.

For purposes of this discussion, acid gas cleanup processes that operate below 420°K (300°F) are defined as low-temperature processes. Processes of this type are widely available, having been used in both the natural gas and chemical process industries:

- Physical Solvent Processes remove acid gases from the raw product gas by physical absorption in an organic solvent. These processes must operate at high pressures since the solubilities of acid gases in the solvents are not sufficiently high at low pressures. Most of the solvents used in these processes have an appreciably higher

affinity for H_2S than for CO_2 and can, therefore, be used in a manner that allows for selective removal of H_2S.

- Chemical Solvent Processes remove acid gases by forming chemical complexes. In most of these processes the solvent is regenerated by thermal decomposition of the chemical complex. These processes are generally identified by the type of solvent used. Amine, ammonia and alkaline salt solutions are the three solvents in common use.

- Combination Chemical/Physical Solvent Processes use a physical solvent together with an alkanolamine chemical solvent additive. The physical solvent absorbs acid gases such as CS_2, mercaptans and COS, which are not easily removed by chemical solvents, while the chemical solvent removes the bulk of the CO_2, H_2S and HCN.

- Direct Conversion Processes produce elemental sulfur from H_2S by oxidation. Some of these processes, such as the Claus and Stretford processes, are not classified as acid gas removal processes in this report; however, they could be used as such. These direct conversion processes are divided into two general categories: dry oxidation and liquid phase oxidation.

- Catalytic Conversion Processes may be divided into two categories: (a) those that convert organic sulfur to H_2S, and (b) those that convert organic sulfur and H_2S to SO_2. Most of these processes are generally not considered to be acid gas removal processes; however, they can be used to convert hard-to-remove acid gases such as COS, CS_2 and mercaptans into compounds such as H_2S and SO_2, which can then be handled by other acid gas removal processes.

- Fixed-Bed Adsorption Processes remove acid gases by adsorption on a fixed sorbent bed. The amount of acid gases removed is dependent on the surface area available for adsorption. Regeneration of the sorbent is accomplished by thermal methods or by chemcial reaction.

Using the criteria previously established for commercialization status, the following acid gas processes were identified to be those which have applicability to this phase of environmental control in the gasification/combined cycle system:

- Physical Solvent Processes:
    - Rectisol
    - Selexol
    - Purisol

- Chemical Solvent Processes:
    - MEA
    - DIPA
    - MDEA
    - DGA
    - Benfield

- Combination Chemical/Physical Solvent Processes:
    - Amisol
    - Sulfinol

A comparison of these is shown in Table 4-5. Clearly, these can be used to meet the environmental standards discussed earlier. However, as seen in their operating conditions, matching problems with gasifiers are apparent especially with regard to pressure. Details of some of these processes are provided in Appendix C2.

The next step in the gas cleanup is to remove the sulfur from the captured streams. These are generally refered to as tail gas cleanup processes. Again, using the commercialization criteria, the following sulfur control processes

TABLE 4-5

COMPARISON OF LOW TEMPERATURE ACID GAS REMOVAL PROCESSES

	CHEMICAL SOLVENT PROCESSES					PHYSICAL SOLVENT PROCESSES			COMBINATION PROCESSES	
	MEA	MDEA	DIPA	DGA	BEUFIELD	RACTINOL	SELEXOL	PURINOL	SULFINOL	AMISOL
Control Effectiveness										
• H_2S	99.9+%	99.9+%	99.9+%	99.9+%	99.9+%	99.9+%	99.9+%	99.9+%	99.9+%	99.9+%
• CO_2	99+%	99+%	DNA	99+%	99.9+%	99.9+%	99.9+%	99.9+%	99+%	99+%
• COS/CS_2	D	DNA	DNA	D	75-99%	99.9+%	99.9+%	99+%	90+%	99+%
• R-SH	D	DNA	DNA	D	68-92%	99.9+%	99.9+%	DNA	DNA	DNA
• HCN	DNA	DNA	DNA	D	99+%	DNA	DNA	DNA	DNA	DNA
• NH_3	DNA	DNA	DNA	DNA	DNA	DNA	DNA	DNA	DNA	DNA
Capable of being Operated Selectively (to remove H_2S without CO_2)	DNA	yes	yes	DNA	yes	yes	yes	yes	yes	DNA
Typical Operating Requirements (per 10^6 scf of gas)										
• Steam (lb)	✓	5,000-10,000	22,000	40,000	16,000-40,000	14,000	3,000	3,000	10,000	✓
• Electricity (kWh)	✓	8-15	85	✓	140-700	300	3,900	300	60	✓
• Cooling Water (gal)	✓	✓	✓	✓	30,000	✓	85,000	13,000	✓	✓
• Fuel Gas						✓	✓	✓		
• Chemicals					✓	✓	✓	✓		
Pressure (psia)	atmospheric	60	15-1,000	not available	100-2,000	300-1,000	300-1,000	1,000	400	200
Temperature (°F)	100-120	80-110	100-140	90-130	280	-30 to -80	20-100	100-110	100-125	90
Discharge Streams Requiring Further Control										
• Gaseous	✓	✓	✓	✓	✓	✓	✓	✓	✓	✓
• Aqueous	✓	NR	NR	✓	✓	✓	NR	✓	NR	NR
• Solid	NR	NR	NR	NR	NR	NR	NR	NR	NR	NR
By-Products	NR	NR	NR	NR	NR	Naphtha	NR	NR	NR	NR
Advantages	Low solvent cost high capacity	Wide range of pressure solvent not degraded	Noncorrosion solvent solvent not degraded	Has absorption of heavy hydrocarbons	Removes organic sulfur and hydrogen cyanide	Good selectivity inexpensive solvent	Low solvent loss	Noncorrosive low solvent loss	Low corrosion solvent not degraded Lower steam requirements than amisol	Easy regeneration of solvent
Disadvantages	Organic sulfur compounds degrade solvent	Corrosion problems greater than MEA Does not remove mercaptans	High pressure needed	Organic sulfur compounds degrade solvent	Incomplete H_2S control without CO_2 control	Low temp. required to limit solvent losses; retains heavy hydrocarbons, high pressure	Retains heavy hydrocarbons, high pressure	Retains heavy hydrocarbons, high pressure	Solvent is expensive	Expensive

NOTES: NR - None reported
DNA - data not available
D - solvent degraden forming nonregenerable compounds
✓ - indicates premenen of a utility requirement or discharge stream

were identified as likely to be available for implementation:

- Primary Sulfur Recovery Processes:
  - Claus
  - Stretford
- Tail Gas Cleanup Processes:
  - Beavon
  - SCOT

Comparison and operating parameters are shown in Table 4-6, with detail process summaries in Appendix C3. We may note that the elemental sulfur end product here is a saleable by-product of the gasification process.

4.5 Matching of Acid Gas Systems to Gasifiers

To develop the kind of considerations that go into the selection and matching of acid gas systems to gasifiers, we provide a brief discussion here. Doing so requires basically a comparison of the data from Tables 2-6, 2-7, 2-8 and 2-9 with that from Tables 4-4, 4-5 and 4-6.

We first observe that any of the six selected gasifiers can use any of the suggested particulate capture processes. Their use is generally independent of inlet conditions, especially pressure.

The next basic step is gas quenching and cooling. Note that the fluidized and entrained bed gasifiers do not require this step. The nature of the gasification process is such that for those configurations, limited, if any, tars and oils are produced. Potential applicability of the several commercial technologies is displayed in Table 4-7. Basically, the matching here involves pressure considerations.

TABLE 4-6

SUMMARY OF SULFUR RECOVERY AND CONTROL PROCESSES

	Sulfur Recovery Process		Tail Gas Cleanup Processes	
	Claus	Stretford	Beavon	SCOT
Development Status	Commercial	Commercial	Commercial	Commercial
Control Effectiveness				
• H_2S	90-95%	99.9+%	99.9+%	99.8+%
• COS/CS_2	90%	-	98+%	98+%
• R-SH	95%	-	DNA	DNA
• HCN	DNA	D	D	DNA
• NH_3	DNA	-	DNA	DNA
• Hydrocarbons	90%	-		-
Operating Requirements				
• Steam		✓	✓	✓
• Electricity		✓	✓	✓
• Cooling Water	✓			
• Fuel Gas			✓	✓
• Chemicals (including catalyst)	✓	✓	✓	
• Process Water		✓	✓	
Discharge Streams Requiring Further Control				
• Gaseous	✓		✓*	
• Aqueous			✓	✓
• Solid	✓			
By-Products				
• Sulfur	✓	✓	✓	
• Other				
Applicability To Coal Gasification				
• Proven				
• Technically Feasible	✓	✓	✓	✓
Disadvantages	High hydrocarbon feed can result in formation of organic sulfur compounds	Does not remove organic sulfur compounds; high pressure		
Advantages	Classical, commercial process Some steam recovered	High turndown Low maintenance	Can use Stretford Steam recovered	Proven system

*If organic sulfur compounds are present in feed stream

D - Solvent degrades forming nonregenerable compounds

DNA - Data not available

✓ - Indicates presence of an operating requirement, discharge stream, by-product, or applicability characteristic

TABLE 4-7

APPLICABILITY OF ENVIRONMENTAL CONTROL EQUIPMENT
GAS QUENCHING AND COOLING PROCESSES

Gasifier	Floccolation Flotation	All-Water Separation	Filtration	Extraction	Absorption	Biological Oxidation	Acid Gas Stripping	Forced Evaporation
Fixed Beds								
Lurgi, Dry Ash	Pressure limited	✓	✓	Pressure limits →				
Wellman-Galusha	✓	✓	✓	✓	✓	✓	✓	✓
Woodall-Duckham	✓	✓	✓	✓	✓	✓	✓	✓
Fluidized Bed								
Winkler	These processes generally not needed since only limited tars and oils produced.							
Entrained Beds								
Koppers-Totzek	These processes generally not needed since no tars and oils produced.							
Texaco	These processes generally not needed since no tars and oils produced.							

Note this ✓ implies process applicability.

For the acid gas removal step, we note that all the gasifiers require such processes. Again, there are severe pressure limits which exclude direct use of some of the technologies. Potential applicability here is shown in Table 4-8. As a general statement, we may conclude that the chemical techniques are largely atmospheric systems while the physical solvent processes are pressurized systems.

We might add here that there is always the possibility of pressurizing the raw gas stream prior to cleanup. This, however, imposes maintainability problems with the compressor system because of the constituents of the raw gas. Thus this step is not recommended.

In concluding, we emphasize that each gasifier manufacturer usually selects a particular cleanup sequence and process specifically applicable to that gasifier. Thus the considerations here should be treated as generic.

4.6 Gas Turbine Combustor and Exhaust Stack Requirements

Finally, we must investigate the nature of the combustion process at the gas turbine, the exhaust gases and the stack cleaning requirements. As may be seen from the discussion of the acid gas cleanup phase, there are sufficient controls on the sulfur process to ensure that the standards cited in Table 4-2 are satisfied. There remains the nitrogen oxides problem.

Current operating experience suggest that nitrogen oxide control is well established, see for example References 9 and 10. Basic systems include wet controls by the use of water injectors or steam injectors. Emerging technologies include a dry system which is somewhat more efficient but does not appear commercially available at present.

TABLE 4-8

APPLICABILITY OF ENVIRONMENTAL CONTROL EQUIPMENT
ACID GAS REMOVAL PROCESSES

GASIFIER	MEA	MDEA	DIPA	DGA	BENFIELD	RECTINOL	SELEXOL	PURISOL	SULFINOL	AMISOL
Fixed bed										
Lurgi, Dry Ash	pressure limited		✓	✓	✓	✓	✓	✓	✓	✓
Wellman-Galusha	✓	✓	0	✓	pressure limited →					
Woodall-Duckham	✓	✓	0	✓	pressure limited →					
Fluidized bed										
Winkler	✓	✓	0	✓	pressure limited →					
Entrained beds										
Koppers-Totzek	✓	✓	0	✓	pressure limited →					
Texaco	pressure limited		✓	✓	✓	✓	✓	✓	✓	✓

Note that ✓ implies process applicability; 0 implies unknown applicability.

4.7 REFERENCES

4.1 Lund, Herbert F., Ed., "Industrial Pollution Control Handbook" McGraw Hill Book Company, New York, 1971.

4.2 Danielson, John A., Ed., "Air Pollution Control Manual" U.S. Environmental Protection Agency, Publication No. AP-40, 1973.

4.3 See any of the "Coal Technology" Conferences collection of papers published in 1978 and 1980.

4.4 Cavanaugh, E.C., et al, "Environmental Assessment Data Base for Low/Medium-Btu Gasification Technology" Volume I, Technical Discussion, U.S. Environmental Protection Agency, EPA-600/7-77-1259, November 1977.

4.5 Cavanaugh, E.C., et al, "Environmental Assessment Data Base for Low/Medium-Btu Gasification Technology" Volume II, Appendices A-F, U.S. Environmental Protection Agency, EPA-600/7-77-125b, November 1977.

4.6 Murin, P., et al, "Environmental Assessment Report: Wellman-Galusha Low-Btu Gasification System" U.S. Environmental Protection Agency, EPA-600/7-80-093, May 1980.

4.7 Mudge, L.K., et al, "Assessment of Environmental Control Technologies for Koppers-Totzek, Winkler and Texaco Coal Gasification Systems" Pacific Northwest Laboratory, Richland, Washington, PNL-3104.

4.8 Robson, F.L. and Blecker, W.A., "Gasification/Combined-Cycle Power Generation: Environmental Assessment of Alternative Systems" Argonne National Laboratory, Argonne, Illinois, ANL/ECT-7, November 1978.

4.9 Federal Register, Vol. 45, No. 240, Thursday, December 11, 1980, pp. 81653-81665, "Denial of Petition to Revise Standards of Performance for New Stationary Sources: Stationary Gas Turbines".

4.10 Federal Register, Vol. 46, No. 72, Wednesday, April 15, 1981, pp. 22005-22006 "Standards of Performance for New Stationary Sources: Stationary Gas Turbines; Proposed Revision of Standard".

5.0 CONVENTIONAL COAL-FIRED COGENERATION CYCLES

In addition to a coal gasification/combined cycle plant, subject of the proceding sections, this study also considers a conventional coal-fired electric and steam central power plant operating in a cogeneration mode. At the expected electric and steam loads of Sewells Point a conventional coal-fired plant would consist of pulverized, stoker-fed or fluidized bed boilers and back pressure or extraction steam turbines. This equipment has long been commercial and, therefore, is readily available from many sources; it has, for many years, been within the engineering state-of-the-art. For this reason discussion of the characteristics of the individual equipment are set forth in less detail than for gasifiers and combined cycles. It is the identification of the most likely optimum combination of boiler and turbine elements and their respective sizing which is the major consideration here.

Considerations for the boiler plant must also include environmental factors that restrict plant design. Plant discharges are required to meet the U.S. Environmental Protection Agency Standards. Of concern here are discharges of sulfur oxides, nitrogen oxides and particulate matter resulting from combustion, as well as coal pile runoff. Other discharges from the power plant that would require treatment are backwash from water treatment and boiler blowoff. The latter are part of current plant operations and installation of new boilers will not significantly alter such operations.

Coal pile runoff can be eliminated from consideration since all coal will be received, stored and transported under cover (see Section 7.0). This also avoids the problem of fugitive dust usually generated by these operations.

The characteristics of coal fired in the plant, and the means of combustion, form the basis of the plant equipment needed to

protect the environment. Coal, when combusted, will require reduction of pollutants. Table 4-1 outlines the performance requirements. Techniques used to meet these standards are also outlined in the following discussion.

5.1 Boilers

From the data in Section 1.0 the existing plant at Sewells Point in 1988 will consist of:

a. Three vintage 1940 boilers, totalling 225,000 lb/hr output, which can only be oil-fired.

b. Four vintage 1942 to 1944 boilers, totalling 415,000 lb/hr of original nameplate output but derated to approximately 360,000 lb/hr. These boilers will have been upgraded by 1985 to be capable of being either pulverized coal-fired or oil-fired.*

c. One vintage 1980 boiler of 200,000 lb/hr output capable of being pulverized coal-fired or oil-fired.

In 1988 the Sewells Point steam plant will also consist of satellite peaking oil-fired plants with 281,000 lb/hr output and a refuse-fired plant of 120,000 lb/hr. The overall steam plant capacity will then be greater than the projected peak steam demand in 1988 by 40%, if all boilers are capable of being fired.

*At the time of completion of this study, the upgrade design and implementation was put on hold by the U.S. Navy. This affects economics and so is addressed in Section 8.0.

From a preliminary analysis of the economic parameters, we have determined that, aside from savings obtainable by increased efficiencies, the major source of benefits accruing to cogeneration that adequately offset the capital investment requirement arise from:

- Substitution of coal for oil and the attendant short and long term savings accompanying this change, and

- Avoidance or diminution of electric costs for both energy and demand charges.

Based on the above equipment two plant configurations naturally arise. One would be the installation of a 225,000 lb/hr boiler operating at 1200 psig and 700°F and the retirement of the three low-pressure oil-fired boilers at Plant P-1. This steam output will serve to define the amount of electric power to be cogenerated.

To produce greater quantities of electric power a second possible configuration would be the installation of three 200,000 lb/hr boilers operating at 1200 psig and 700°F. In such a scheme not only the three existing oil-fired boilers but also the four then recently upgraded, but still quite old, coal-fired boilers would be retired. Their cost to upgrade will have been fully paid back by 1988. This scheme would provide more electric cogeneration and, therefore, greater economic benefits.

The boilers under consideration would be stoker-fired or pulverized coal-fired boilers with flue gas desulfurization and fluidized bed combustion boilers. The essential characteristics of these follow recognizing that New Source Performance Standards are applicable.

5.1.1 Stoker-Fired Boilers With Flue Gas Desulfurization

Industrial stoker boilers can be produced as top or bottom supported types. A top hung boiler uses two drums to connect convection surface at the boiler outlet and permits factory fabricated water wall surface to be used as the side enclosure. Thus, the boiler can be structurally sound and provide a minimum of air or gas leakage.

The use of tall vertical walls also improves the ability of the boiler to remain free of slag. Wall deslagging units are used in conjunction with retractable soot blowers in the convection pass. A superheater is generally set at the entrance to the convection boiler bank. A typical cross-section is shown in Exhibit 5-1.

To improve boiler efficiency, an economizer surface would be set in the downward pass of the flue gas as it travels towards the induced draft fan. This surface preheats the boiler feedwater and reduces flue gas temperature.

The power plant will house a water treatment system, deaerating and closed feedwater heating system, compressed air system and fuel transfer systems. The major components in altering the power plant operation include:

- Coal bunker with under-bunker transfer conveyor.
- Coal scales.
- Coal chute with distribution cone to prevent segregation.
- Coal spreader with feed assembly.
- Traveling grate with siftings hopper.
- Ash hopper at boiler front with clinker grinder and ash removal air lock for dry bottom ash removal.
- Boiler with superheater element and downpass economizer.

STOKER-FIRED POWER PLANT
SECTION

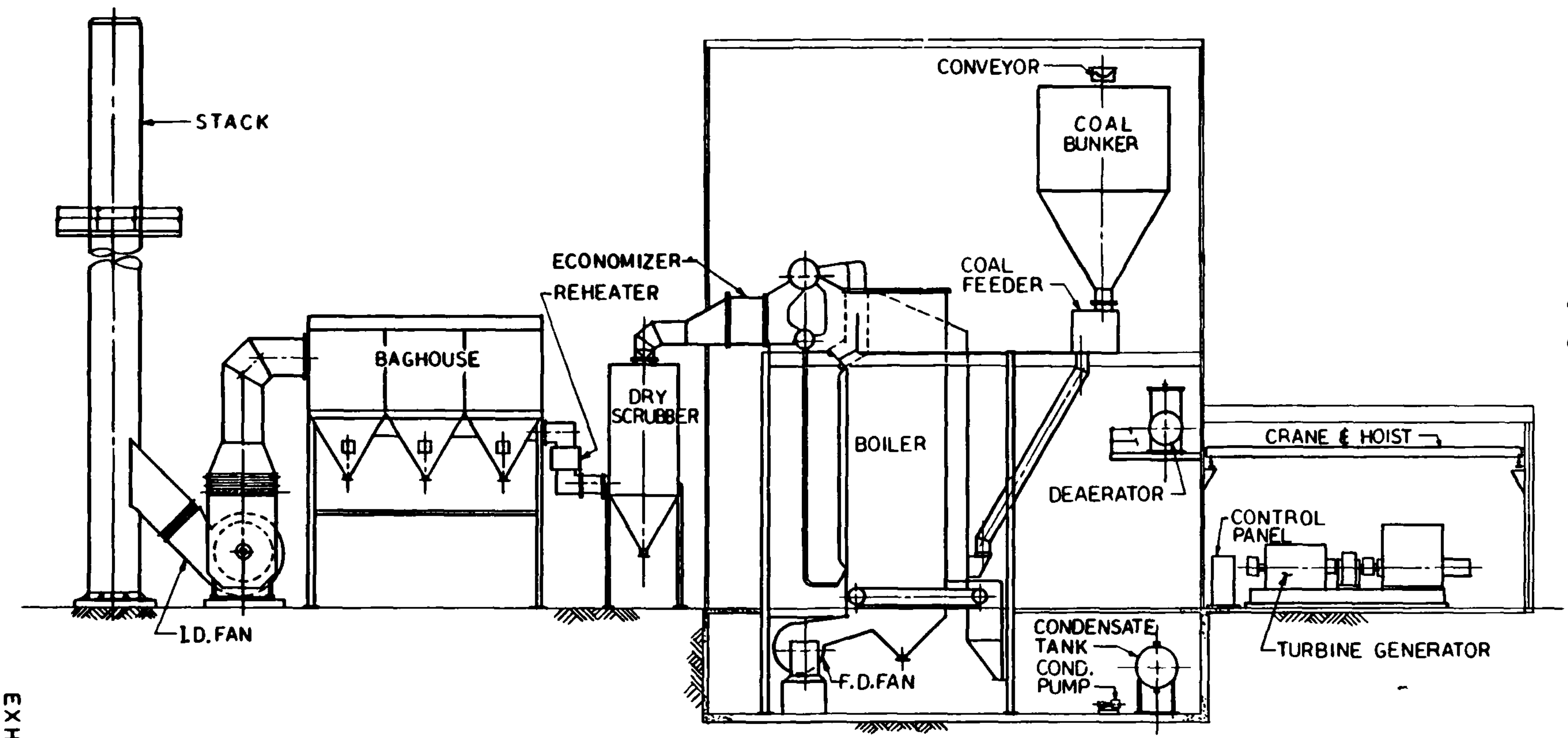


EXHIBIT 5-1

- Mechanical dust collector with fly ash reinjection nozzles.
- Forced draft fan.
- Ductwork leading to a final air pollution control element.
- Induced draft fan.

Ash removal equipment would accept ash from the siftings hopper, ash hopper air lock, duct low points and final air pollution control element. Ash would be pneumatically conveyed to storage silos.

Stacks will be designed for combustion gas flows at approximately 110 feet per second. Stacks will be insulated, double-wall type to prevent condensation and acid attack.

The new boilers will be arranged for full firing of fuel oil as well as coal. Oil burners will be set in one side or the rear wall and take supply from the existing fuel oil system via a day tank.

Stoker-fired boilers, without flue gas cleanup, discharge sulfur oxides and particulate matter in excess of regulation limits. Although stokers generate less fly ash than pulverized coal burners, discharges are still in excess of permissable emissions. Further, cleanup cannot be met by the installation of mechanical dust collectors alone. This study recommends, therefore, the installation of a system that permits both pollutant captures in one system. During the preliminary design phase, a determination of the actual system, wet or dry type, will be made.

The choice is a matter of economics, after consideration of other items such as reliability, availability, longevity and

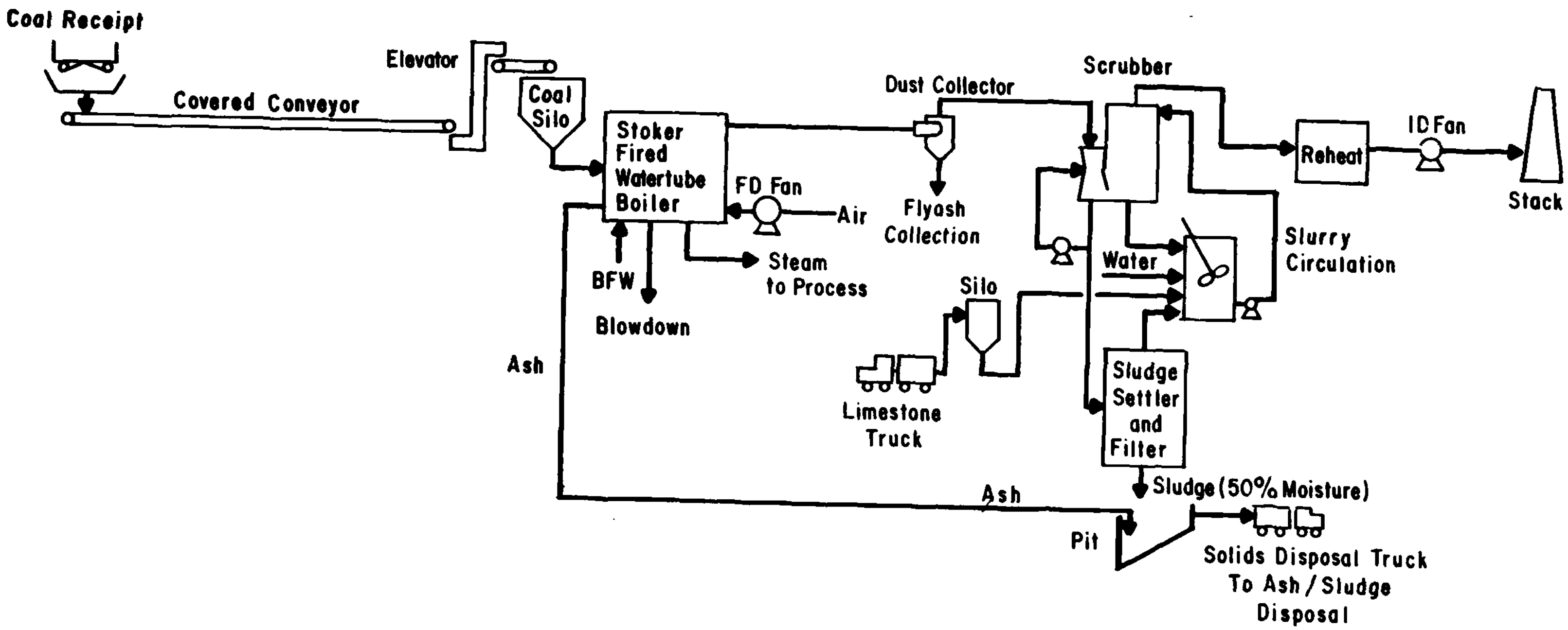
ease of maintenance. A typical schematic of an FGD system is shown in Exhibit 5-2 for stoker-fired coal.

The wet type includes a mixing chamber, usually a venturi nozzle that permits intimate contact between the gas and a liquid bath, and a combination contact tower (scrubber) and liquid removal chamber. Particulate matter is carried along with the gas stream, making contact with the chemically treated liquid and is captured with the chemical reaction precipitates formed in capture of the SO_2 gas. The dry type includes a spray chamber in which the flue gas is sprayed with a lime solution, an SO_2 sorbent. Particles and the result of the chemical reaction between SO_2 and the sorbent are then trapped on filter media in a baghouse. The induced draft fan is downstream of the baghouse and thus sees clean air at a temperature of approximately 150°F.

Both the wet and dry methods require that the flue gas be reheated after treatment. Heat is added to permit the gas to form an acceptable plume. On a cold, dry day moisture in the flue gas would rapidly condense and fall as rain in the immediate area or as ice crystals in extreme cold. Heat can be taken from the boiler in the form of a steam coil in the discharge of the stack, or can be taken from the flue gas. In the latter case, 80% clean air is taken from a point ahead of the scrubbing or spraying process and mixed with the cleaned air to increase its temperature. Precise measurement of particulate matter and SO_2 concentration downstream of the process would dictate the quantities of untreated gas that could be added.

Equipment necessary for the wet systems vary depending upon the system selected. Either a lime or combination lime and soda ash sorbent would be stored in silos. Dry chemicals are mixed with recycle water and are then pumped to the venturi section, to the scrubber tower and in some processes, to a

CONVENTIONAL BOILER SYSTEM FIRED WITH HIGH SULFUR COAL



BOILER TYPE	Watertube, spreader stoker fired, field assembled
ENVIRONMENTAL PROVISIONS:	
SO_2	Meets criteria
NO_x	Meets criteria
Particulates	Multiclone dust collector for bulk fly ash; scrubber for final cleanup

POPE, EVANS AND ROBBINS

EXHIBIT 5-2

thickener. In a wet system the scrubber tower performs two functions, completing the chemical reaction to fix SO_2 in a precipitable form and removing excess solution from the off-gas. The dewatering feature becomes an important function in the matter of forming fog in the immediate area. There is a need for various mixing tanks, balancing tanks and recycle tanks to provide for small disturbances in the system flow. However, one item that occupies the most space is a thickener. The thickener is needed to facilitate waste separation from recyclable solution. It permits a more concentrated solution to be pumped to a final filtering station, generally a vacuum filter.

Each boiler would be served by a scrubber with one common thickener. Under-flow from the thickener would discharge to the filter house comprising three filters, one of which is standby, three vacuum pumps and elevated tankage needed for chemical feeds and conveying equipment. The vacuum filters produce a sludge, approximately 50 percent dry, as compared to the 3 or 4 percent solution that is fed to them. The sludge is an inert cake comprising the chemically combined sulfur with fly ash. Sludge is delivered to silos having two wedge shaped compartments, each with multiple discharge screws to drop sludge into lugger pans. These are lifted onto trucks for disposal. Because there is 50 percent moisture in storage, the bottoms of the silo are heated to assure removal during cold weather.

Draft losses associated with the venturi and scrubber tower must be overcome. For the purpose of this study, we have assumed installation of a new induced fan for each new boiler that would assume draft loss between furnance and stack exit. A heater section would be installed to permit the gases to be dried before discharge. The matter of plume rise would be considered during an environmental analysis of the installation.

In the same manner as a wet scrubber, a dry system could be installed. This would include a spray chamber, in which sorbent is finely divided and dispersed in the flue gas stream, and a baghouse to act as a final filtering element. The flue gas temperature is lowered by the evaporation of the sorbent chemicals, however, the dew point is not reached and the gas entering the baghouse is not saturated with water vapor. Fly ash deposited on the bags, together with reacted sorbent, serve to trap additional material. The bags are puffed clean when the pressure drop through the baghouse exceeds its loss limit, approximately 7-inches W.G., and the dust falls into hoppers for collection. The cleaned gas is then reheated to assure buoyancy and discharged to the atmosphere through an induced draft fan. As in the case of the wet FGD system, there are chemical feeds required with associated tankage, mixers and pumps. Spray dryers are necessarily large vessels to permit the residence time required for reaction. Baghouses are also large to accommodate the volume of gas while maintaining a reasonable velocity through the fabric filters.

Selection of a wet or dry scrubber depends on life cycle costing, once technical factors have been addressed. Particulate matter removal from the flue gas stream is required when firing a spreader stoker; nitrogen oxides are probably within limits because of low firebox temperatures; but, sulfur oxide removal will depend upon the coal source being fired. Current air pollution regulations permit firing of 12,000 Btu/lb coal without removal of sulfur oxides from flue gas if SO_2 emissions are comparable to a sulfur content of 0.7%. This study assumes installation of a dry scrubber to permit reduction of sulfur oxides while meeting particulate matter standards as well.

5.1.2 Pulverized Coal Fired Boiler With Flue Gas Desulfurization

High pressure boilers in the size required at Sewells Point are readily available for pulverized coal (PC) firing. PC firing has the advantage of improved efficiency over stoker firing, but has generally higher cost; selection would be based on life cycle analysis. PC boilers are generally top hung, making use of water cooled walls to absorb radiant energy and either radiant or convection superheaters. Air preheaters are used in lieu of economizers to reduce exit gas losses. Some combustion air, after warming in the preheater, passes through the coal pulverizers and acts as the conveying medium for the pulverized coal. The remainder of the air supply enters the windbox and supports combustion. Heated air increases furnace temperature and thus the chance for nitrogen oxide formation increases, but control is possible.

The other pollutants of sulfur oxides and particulate matter require attenuation in the same manner as in stoker firing. Compliance coal may be used in lieu of desulfurization, but is not likely to be procurable for a plant of this magnitude. The quantities of fly ash generated by PC firing are greater than those in stoker firing, but the principles of collection and treatment of pollutants is essentially the same as for the stoker-fired cased previously described.

PC boilers use oil firing for light off and are readily sized for 100% use of oil in case of an inability to deliver coal to the power plant.

5.1.3 Fluidized Bed Combustion Boilers

A potential alternative to the stoker fired boiler is an atmospheric pressure fluidized bed combustion (FBC) boiler. In FBC desulfurization occurs during combustion and

the flue gas needs no further sulfur capture. Nitrogren oxide formation is not a major concern as furnace temperatures are less than in stoker-fired units. However, particulate cleanup is still required. See Exhibit 5-3 for an overall schematic of an FBC system.

Fluidization is developed to provide a violently bubbling bed of relatively large particles of coal and limestone, insuring good process control for combustion and sulfur removal. Coal is ignited in the bed and sulfur oxide is captured as expressed by the following equations:

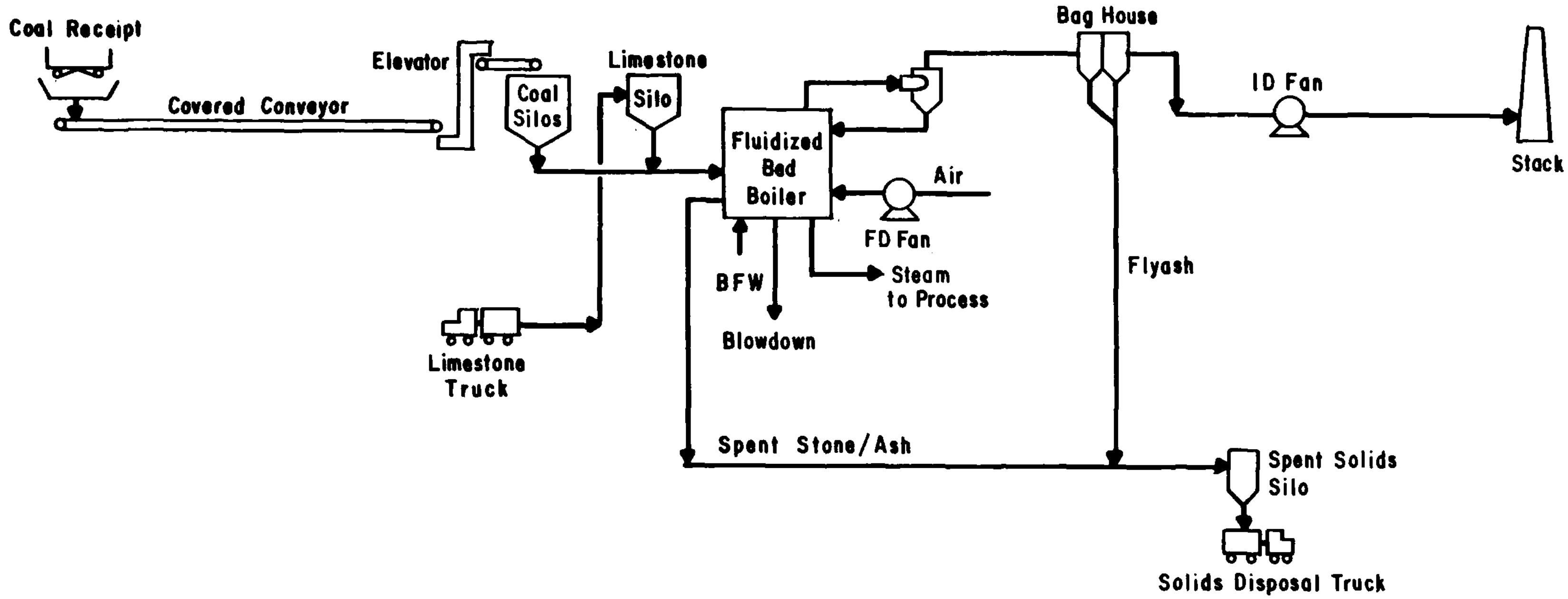
$$CaCO_3 + Heat = CaO + CO_2$$
$$CaO + SO_2 + 1/2\ O_2 = CaSO_4$$

The fluid bed operates at 1500°F, at which temperature sulfur is most readily captured. A typical schematic design for an FBC boiler is shown in Exhibit 5-4 and a FBC installation in Exhibit 5-5.

Since this technology is relatively new, some additional discussion is warranted. The development of fluidized bed combustion on a commercial scale is an essential element of programs of the U.S. Department of Energy and others. The FBC technology closest to commercial readiness is atmospheric fluidized bed combustion, as currently being demonstrated at Georgetown University. The advantages of FBC over traditional coal combustion are:

- Low SO_2 emissions without the use of sulfur scrubbers. This is accomplished by burning the coal in a fluidized bed with sorbent material (lime, limestone, or dolomite). The sorbent material reacts with SO_2 to form calcium sulfate, thus removing it from the flue gas.

FLUIDIZED BED BOILER SYSTEM FIRED WITH HIGH SULFUR COAL

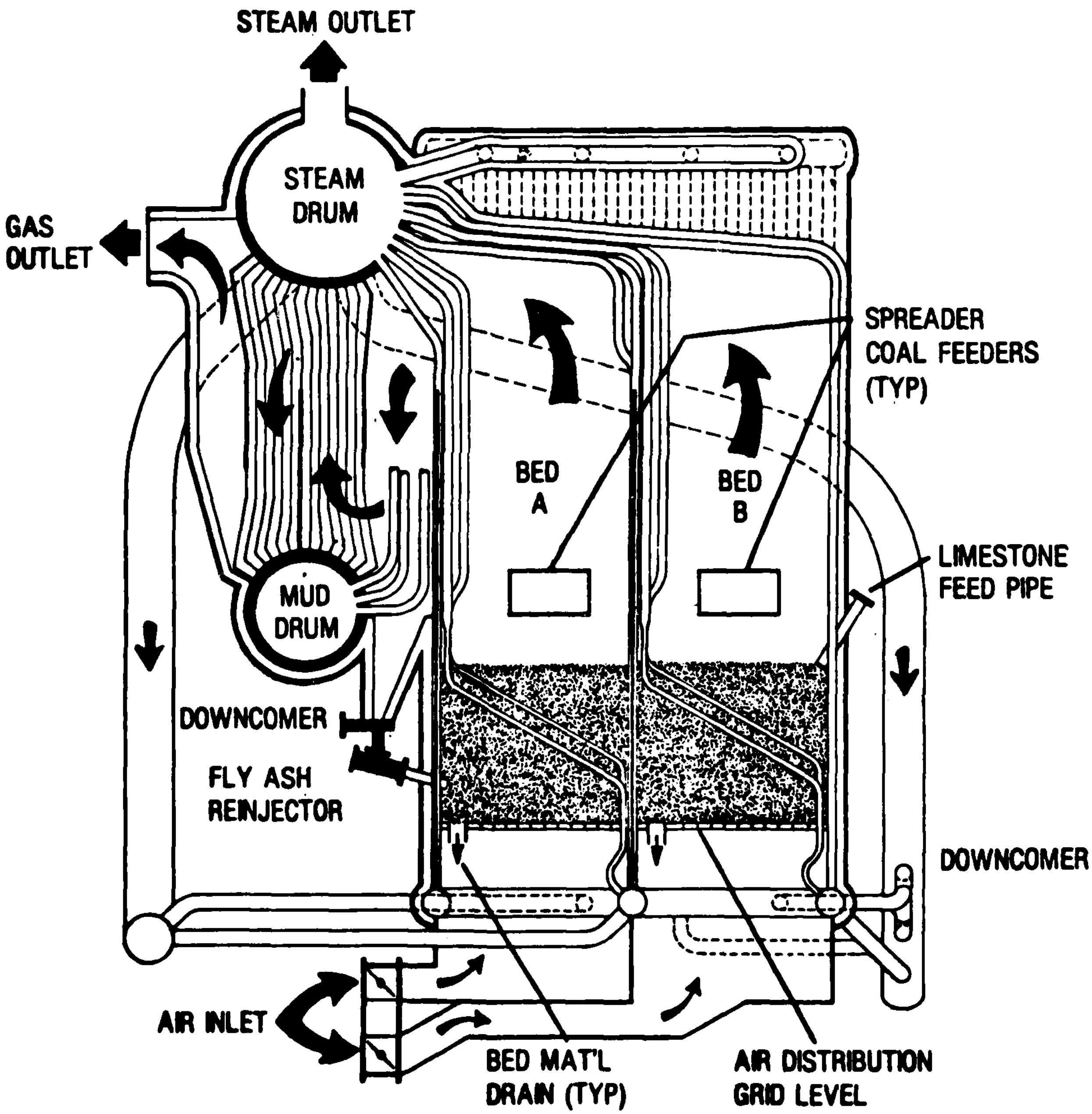


BOILER TYPE	Fluidized bed, watertube, atmospheric pressure
ENVIRONMENTAL PROVISIONS:	
SO_2	Meets criteria
NO_x	Meets criteria
Particulates	Bag House

POPE, EVANS AND ROBBINS

EXHIBIT 5-3

SCHEMATIC DESIGN
FLUIDIZED BED BOILER



FBC BOILER SYSTEM

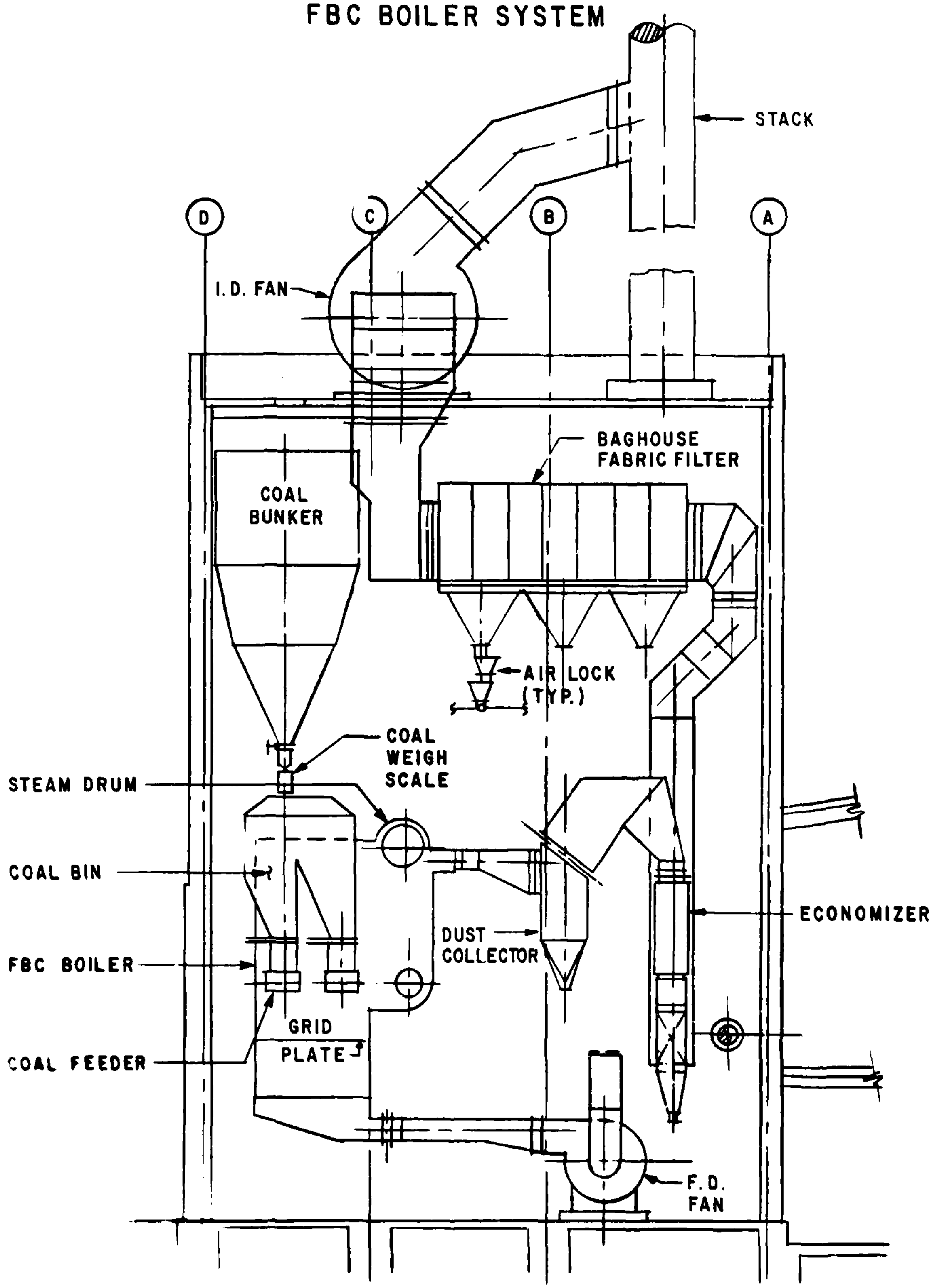


EXHIBIT 5-5

- Lower NO_x emissions than conventional boilers due to FBC's lower reaction temperature (1500-1600°F versus 2000°F).

- Reduced ash agglomeration or slagging due to the lower combustion temperature.

- High heat transfer rates (5 to 8 times convection heat transfer rates) from placing tubes in the fluidized bed.

The efficiency of an FBC is approximately the same as a stoker fired boiler with a grate. Coal usage is therefore the same and coal may be introduced in the same manner as in a standard stoker unit. Some suppliers provide pneumatic injection.

Limestone, the sulfur dioxide sorbent, is a sized dry product that can be transported via pneumatic conveyor. It is stored in silos adjacent to the boiler room with a close-by storage at each boiler.

As calcium sulfate and ash accumulate above the boiler grid plate, the bed height must be reduced to an optimum operating level to permit introduction of fresh limestone sorbent. At Georgetown this is accomplished by means of a water cooled screw. The cooled bed material is then pneumatically conveyed to a storage silo prior to disposal.

Flue gas leaving the boiler passes through a cyclone type mechanical collector. Particles greater than 15 microns diameter are returned to the bed to reduce carbon loss. The gas stream including smaller particulate, is passed through an economizer where the gas temperature is reduced to 400°F, and then through a particulate capture system. Such systems trap more than 99% of the influent particulate. Material trapped

is pneumatically conveyed to a silo for storage prior to disposal.

The furnace operates at a balanced draft, that is, a point over the bed will be at atmospheric pressure. The pressure beneath the grid will be approximately 60-inches water gauge. Two fans are used: a forced draft fan to produce the pressure beneath the grid which fluidizes the bed of material, and an induced draft fan to overcome the losses encountered in moving the flue gas through the particulate collection areas and then through the stack to atmosphere.

Fans use inlet vane control to reduce motor power under low loads, however, one or more fans could be turbine driven depending upon the steam balance of the plant.

Other auxiliary equipment normally associated with a boiler plant are the same for an FBC as they are for other forms of firing. These would include chemical treatment, continuous blowdown, combustion control and light-off system with burner management system.

Note that an FBC boiler has a turndown ratio of 4 to 1. This is accomplished by operating with two parallel beds within one boiler. One bed receives recirculated material and acts as the lead bed; the other bed operates only when the boiler operates at more than 50% load. Each of the beds has 2 to 1 turndown ability and thus each boiler can operate between 25 and 100%.

Consider finally the particulate control options for FBC. The particulate size entering the final filter element is small. A cyclone collector ahead of the final filter will pass particles smaller than 15 microns, approximately 10% of particulates. Two choices can then be made:

- Baghouse (fabric filter)
- Electrostatic Precipitator (ESP)

Unless there are unusual duct arrangement difficulties, pressure loss from inlet to outlet of an ESP is less than 1-inch of water. On the other hand, a baghouse will generally operate in the 6 to 8 inch loss range. This appears as an increase in motor horsepower required for the fan system. Provision for oil firing is required as oil soot can create plugging of collection hoppers. As a mechanical collector will probably reduce oil soot below particulate matter emission standards, a baghouse may be by-passed during an oil firing regime. This would also be true on start-up if oil is the startup fuel.

Selection of a baghouse or an ESP to work in conjunction with an FBC boiler is a matter for life cycle costing, once technical requirements have been met.

5.2 Steam Turbine

The steam turbine can be either back pressure or extraction-condensing type. For the conditions here, a back pressure turbine would produce on the order of 1 kW for each 43 lb/hr of steam flow. The extraction-condensing turbine can produce electric power in a wide range regardless of export steam flow. The consideration of electric energy charge avoidance and reduction of demand charges especially in schedules with a ratchet clause bear on the selection between back pressure and extraction-condensing turbines.

In parallel with the outputs shown in Section 3.0 for the combined cycle, we next provide data for the performance of the back pressure mode of operation. The basic assumptions and overall results are shown in Table 5-1 for both options on sizing discussed earlier, i.e., one 225,000 lb/hr or three 200,000 lb/hr boilers.

TABLE 5-1

SUMMARY OF RESULTS FOR BACK PRESSURE TURBINE OPERATION

ASSUMPTIONS:

- Boiler Efficiency at 80%
- Boiler Feedwater Intlet at 353.5 Btu/lb, t = 380°F
- Turbine Throttle at 1200 psig/700°F
- Turbine Efficiency at 72%
- Turbine Back Pressure at 340 psig
- Unit Availability is 100%

Size (lb/hr)	No. Of Boilers	Turbine Size (kW)	Power Produced (10^6 kWh/yr)	Steam Produced (10^9 lb/yr)	Steam Export (10^9 lb/yr)	Coal Consumption (tons/yr)
225,000	1	5,000	41.5	2.0	1.4	95,300
200,000	3	13,000	88.4	4.2	3.0	203,100

Results do not account for down time or reliability. They should be used as preliminary estimates of power produced, steam produced and coal consumption. Steam exported accounts for internal plant use for feedwater heating as compared to steam produced.

6.0 UTILITY INTERFACE

As recently as September 1979, Virginia Electric Power Company (VEPCO), the electric utility supplying Sewells Point Naval Complex, was suggesting a "dump" power credit of 0.005/ kWh for energy generated at a Navy facility in excess of Navy requirements and introduced into the utility grid. In the very last few years the climate for utility encouragement of cogeneration has changed dramatically. Public Utilities Regulatory Policies Act of 1978, PURPA, has been the cause of this change. It has resulted in VEPCO recently issuing a Cogeneration Schedule.

6.1 Utility Attitudes Toward Cogeneration

The utility attitute toward cogeneration, within its financial and economic requirements, is one of positive cooperation in the implementation of either of the following cogeneration schemes:

- Selective Energy - wherein the cogenration project is sized to provide only a portion of the SPNC electric and steam requirements and reliance is placed on the utility to provide the remaining site electric requirements and on the existing Navy boilers to provide the remaining site steam requirements.

- Total Energy - wherein a larger cogeneration project is sized to provide all the SPNC electric and steam requirements and excess generated is distributed by the utility.

The utility attitude of cooperation extends to Navy, utility or third-party ownership and/or operation and maintenance and to consideration of any joint venture thereof.

6.2 Effect of Presently-Issued Rate Schedules

SPNC presently pays for its electric consumption under "Schedule MS - Federal Government Installation", effective 7-1-80. A copy of this Schedule is incuded herein in its entirety in Appendix D; the features of most interest are:

- The kW demand charge of ¢6.22 per kW has an eleven month ratchet, which results in essentially year-round payment of 90% of the highest peak encountered during the year.

- While the energy charge is stated as 1.546 cents per kWh, the effect of the fuel adjustment clause has resulted lately in a charge of about 3 cents per kWh.

VEPCO's cogeneration schedule is set forth in its "Schedule 10 - Cogeneration and Small Power Producer Service", filed 12-23-80, a copy of which is also included in Appendix D. Important features are:

- the energy and demand charges for purchase from and sales to the utility vary both with time of day and time of year.

- the energy charge for sales to the utility are 93% to 96% of the charge for purchase from the utility.

- energy purchase charges vary from 2.760 to 4.429 cents per kWh; energy sales charges vary from 2.560 to 4.229 cents per kWh.

- there is no fuel adjustment clause.

- demand charges vary from $1.69 to $1.95 per kW but, importantly, there is no eleven-month ratchet.

- there is a distribution demand charge which does contain an eleven-month ratchet; however, this charge is only $0.87 per kW.

VEPCO advised that its recent cogeneration Schedule 10 was derived from consideration of marginal costs while its existing Schedule MS has historically been developed by consideration of embedded costs. It intends to update Schedule MS derived on the basis of marginal costs which can then be expected to raise the MS costs. VEPCO projections are that, on a differential escalation basis, it expects the MS prices to fall 0.3 percent between 1980 and 1985 which is consistent with current Department of Energy projections (Federal Register, October 27, 1980).

VEPCO also intends to update the cogeneration Schedule 10 from time to time as experience demands. For example, by 1985, VEPCO expects to have equalized summer-winter annual peaks and the time-of-year aspects of Schedule 10 prices will have to be revised accordingly.

VEPCO has generally not been interested in selling steam because of concern about the quality of the condensate return and consequently does not have a steam sale schedule. Further, aside from the substantial SPNC steam load, there is only a minor additional nearby residential load with the next possible industrial steam load some 10 miles away. Wheeling steam from SPNC, therefore, does not seem possible.

6.3 Utility Policy for Ownership, Operation and Maintenance

VEPCO is interested in small and large cogeneration projects in order to avoid future generation additions requirements.

It is also interested in ownership and/or joint venturing in any project which would bear a competitive capital requirement

less than what they would expect to have to make for future required system additions. VEPCO's description of its future needs suggests:

- In 1985, its second 500 kV line will be operational and it will have a very firm system.

- In the 1990's, it would be looking to adding generation blocks of two 500 MW units. It is turning away from 1000 MW units because of reliability and because economy of scale seems to be disappearing.

- VEPCO expects that the 1980 expansion costs are in the $1000 to $1200/kW range. It offered no thoughts of future escalation; at a nominal rate of 7 percent per year, the price would be in the $1700 to $2100/kW range in 1988.

- Any scheme which would assist VEPCO in its financing would be of particular interest: such as U.S. Government payment of the differential between the required capital requirement and the price VEPCO would normally expect to have to pay.

- The Norfolk area, with its load density, would be a prime candidate for a jointly-owned project.

Due to the absence of nearby steam demand other than SPNC, sizing of a cogeneration project at say 200 or 250 MW would result in steam generation greatly in excess of SPNC requirements. This excess steam can be used through standby steam turbines but whether this can be done competively with utility production is questionable.

6.4 Utility Energy Source Displacement

VEPCO's 1980 energy source mix is 37% nuclear, 23% oil and 40% coal; with the completion of Anna Point III, the mix will be 48% nuclear, 15% oil and 37% coal. VEPCO suggests that these latter values be used for consideration of the displaced utility energy.

6.5 Constraints

VEPCO know of no laws, regulations, policies, agreements, etc., which either constrain a cogeneration option to be considered or must be considered in the design.

6.6 Power Cost Reduction Potential

There is a potential for sizable power cost reduction under both Schedule MS and cogeneration Schedule 10, for either a coal gasification/combined cycle cogeneration scheme or, to an appreciably lesser degree, a conventional coal-fired cogeneration plant.

As set forth in Section 1.0, SPNC 1988 steam and electric requirements suggest that there is a coincident year-round demand of 50 to 60 MW electric and 270,000 lb/hr steam. A selective energy cogeneration project sized at this base load will result in a substantial decrease in electric costs due to avoidance of energy charges under Schedule MS; because of the ratchet, avoidance of demand charges could not be realistically expected without the addition of uneconomical standby equipment to provide the required reliability. The electric savings to be gained by cogenerating 60 MW electric of the 1988 SPNC load are in the order of $13,900,000 under the MS schedule.

A similar base load project will result in even greater decrease in electric costs by use of the cogeneration Schedule 10. This is due to the ability to sell back to the utility

and, also, because of lack of a ratchet clause, the avoidance of demand charges for nine or ten months of the year is feasible. The electric savings achieved by using the cogeneration Schedule 10 for the same load would be in the order of $6,300,000 greater than those achieved by using the existing Schedule MS.

Sizing the cogeneration project at greater than the above base load point, say at 70 MW, results in increased electric savings of a magnitude not sufficient to offset the required increase in project coal input. This supports the idea that the SPNC should not replace the public utility.

Sizing the project to a 50 MW base load point results in decreased electric savings but these are more than offset by the decreased coal fuel input required. For this reason, the life cycle analyses of the coal gasification/combined cycle cogeneration in Section 8.0 are based on a base load of 50 MW.

Table 6-1 illustrates the above. Calculations for Table 6-1 are included in Section D of the Appendix.

Section IV of the cogeneration Schedule 10 gives the option of purchasing the entire SPNC electric requirement under the existing Schedule MS and selling the entire cogeneration project electric output to the utility under cogeneration Schedule 10. The total yearly bill under this option would be:

Purchase Electric Requirement - Schedule MS	
Demand	$ 8,722
Energy	16,023
Sell Project Output - Schedule 10	
Energy	-13,496
Total	$11,249

TABLE 6-1

COMPARISON OF ELECTRIC SAVINGS BY COGENERATING

($000)

	SCHEDULE MS		SCHEDULE 10		
	No Cogen	Cogen 60 MW	Cogen 60 MW	Cogen 70 MW	Cogen 50 MW
'79 Bill - '88 Load					
Demand - kW	8,722	8,722	1,244	1,088	1,398
Demand - Distribution	-	-	1,337	1,337	1,337
Energy	16,023	2,131	2,000	466	4,879
Total	24,745	10,853	4,581	2,891	7,614
Refer to App. Sh.:	(1)	(1)	(1)	(8)	(8)
Savings Cogen 60 MW - Schedule MS:		13,982			
Increased Savings Sched. 10 vs. MS Cogen 60 MW:			6,272		
Increased Savings Cogen 70 MW vs. 60 MW - Sched. 10:				1,690	
LESS: Increased Coal (62,900 tons x $55 ton)				-3,460	
TOTAL (Loss):				-1,770	
Decreased Savings Cogen 50 MW vs. 60 MW - Sched. 10:					-3,032
PLUS: Decreased Coal					3,460
TOTAL (Gain):					428

This $11,249,000 yearly electric bill compares with the yearly bill of $7,614,000 using cogeneration Schedule 10 in its entirety, confirming that the purchase-sell option should not be exercised.

7.0 SITE CONSIDERATIONS

In this section we provide some preliminary remarks for siting the coal gasification/combined cycle power plant and the attendant coal storage for that cycle or for the conventional steam power plant system addressed in Section 5.0. The critical feature here is coal storage, and it is the major focus of this discussion. Other remarks are also offered regarding overall system location.

A coal gasification/combined cycle plant, of the selective energy type envisioned here, indicates that optimum sizing would be at 60 MW with 290,000 lb/hr steam recovery. This calls for essentially base load, full-time operation. The cycle performance developed in Section 3.0 indicates that the coal feed requirements would be approximately 1000 tons/day for such a plant. The area occupied would be as shown in Exhibit 7-1. Elements of the plant are shown in Exhibit 7-2 and are discussed below.

In accordance with NAVFAC DM-3 criteria the minimum 30 day storage requirement is at 100 percent load factor. The design permits storage of 30,000 tons of coal, meeting the criteria, as the source of coal is nearby with a mainline railroad system linking the plant to the producers. It should be noted that a No. 6 fuel oil supply is available to be fired in the conventional boilers remaining in Building P-1.

Coal storage will comprise a large fraction of project site requirements and it is important to identify it. To assure covered storage in this volume, an automated stacking and reclaiming system is chosen. Ground water is near the surface in this area and a below-grade relcaim system should be avoided. The selected system of coal receipt includes:

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EVALUATION OF COAL GASIFICATION/COMBINED CYCLE POWER PLANT FEAS--ETC(U)
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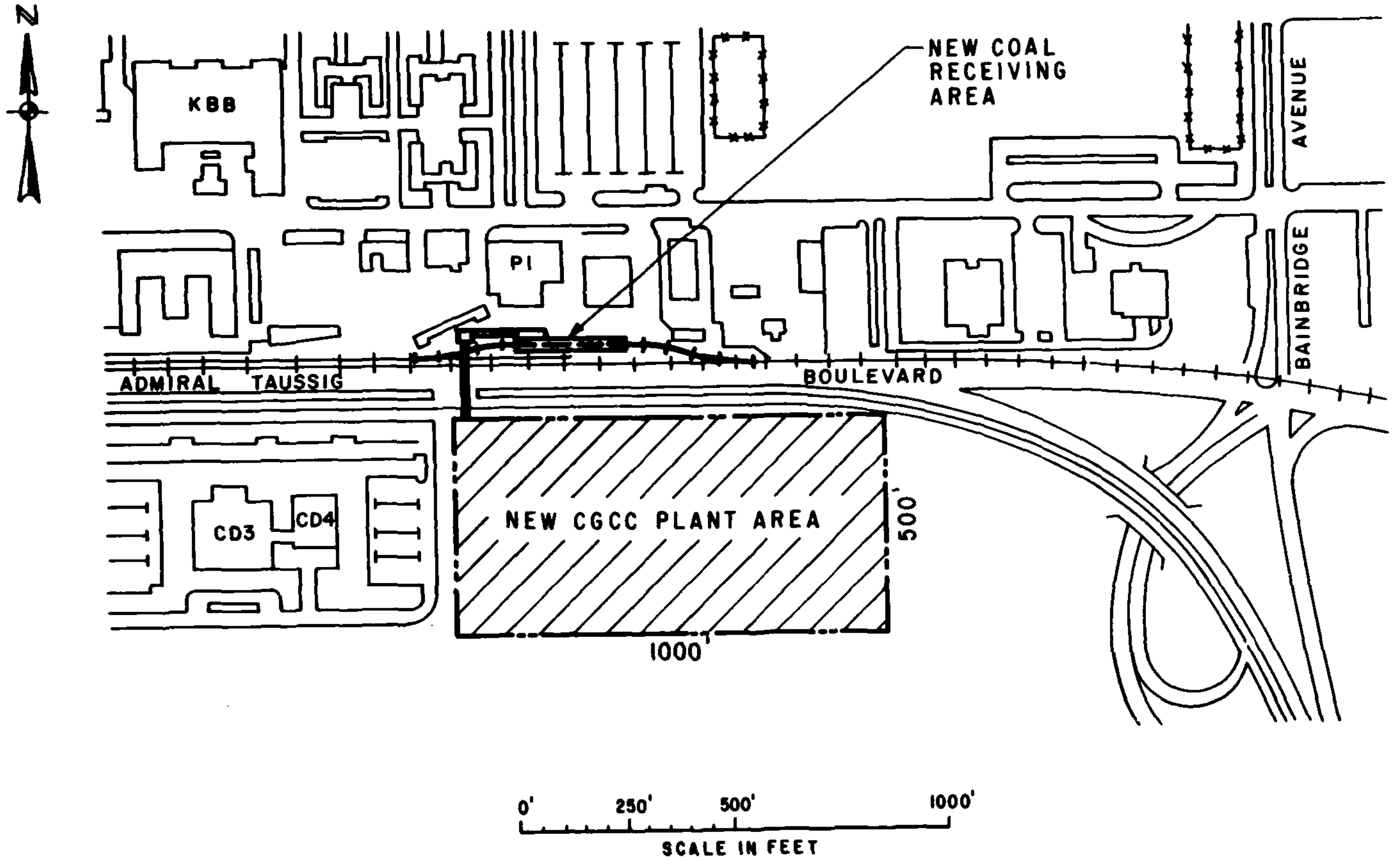
NEW COAL GASIFICATION/COMBINED CYCLE PLANT LOCATION



POPE, EVANS AND ROBBINS

EXHIBIT 7-1

SITE PLAN

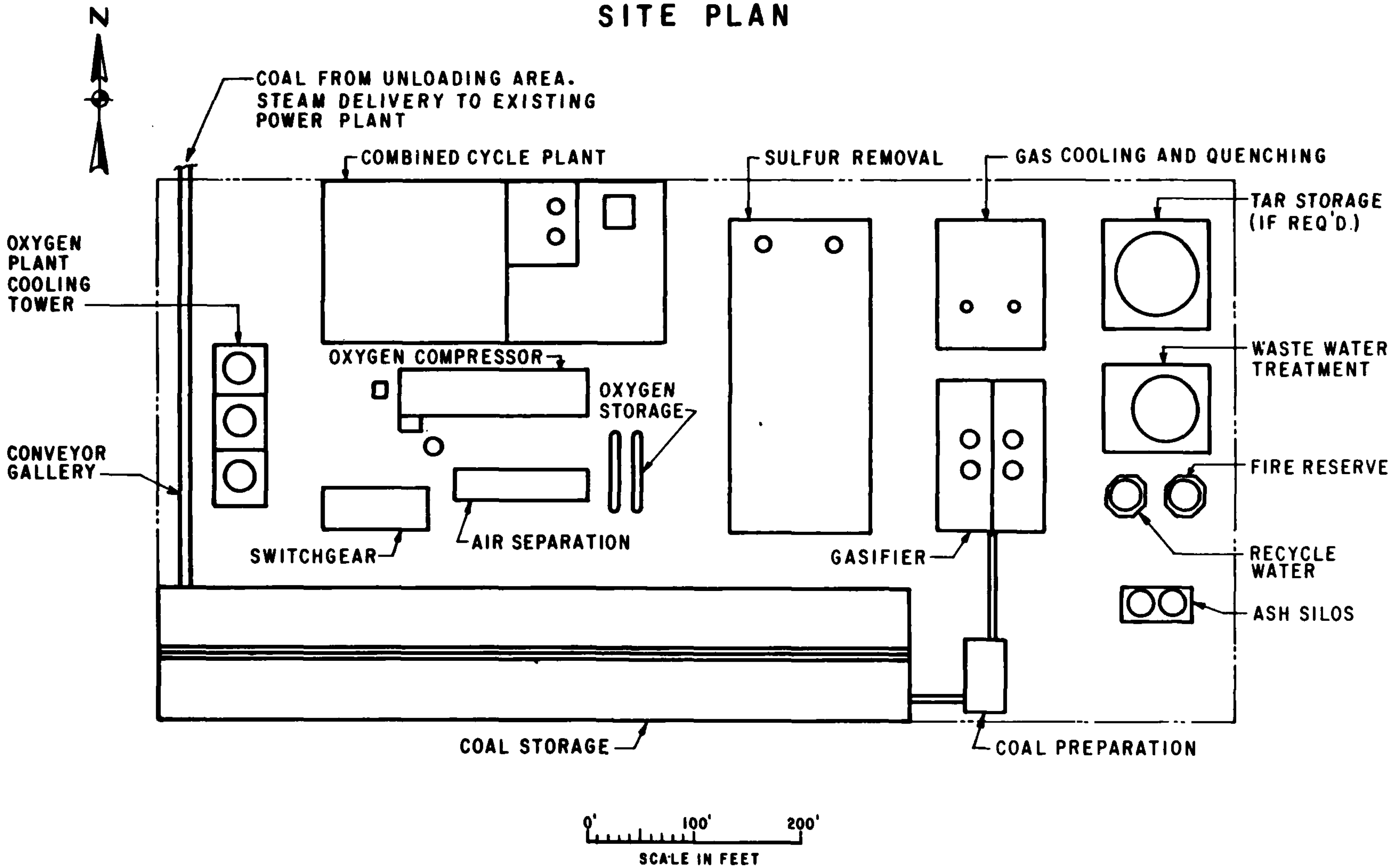


EXHIBIT 7-2

POPE, EVANS AND ROBBINS

- Rail siding with thaw shed, car discharger and dust suppression system.

- Four coal hoppers with feeders to accept discharged coal.

- Dual skip hoists to unload the hoppers at the rate of 250 ton/hour.

- Conveyor from the rail siding and skip hoist tower across Admiral Taussig Boulevard to the coal storage building.

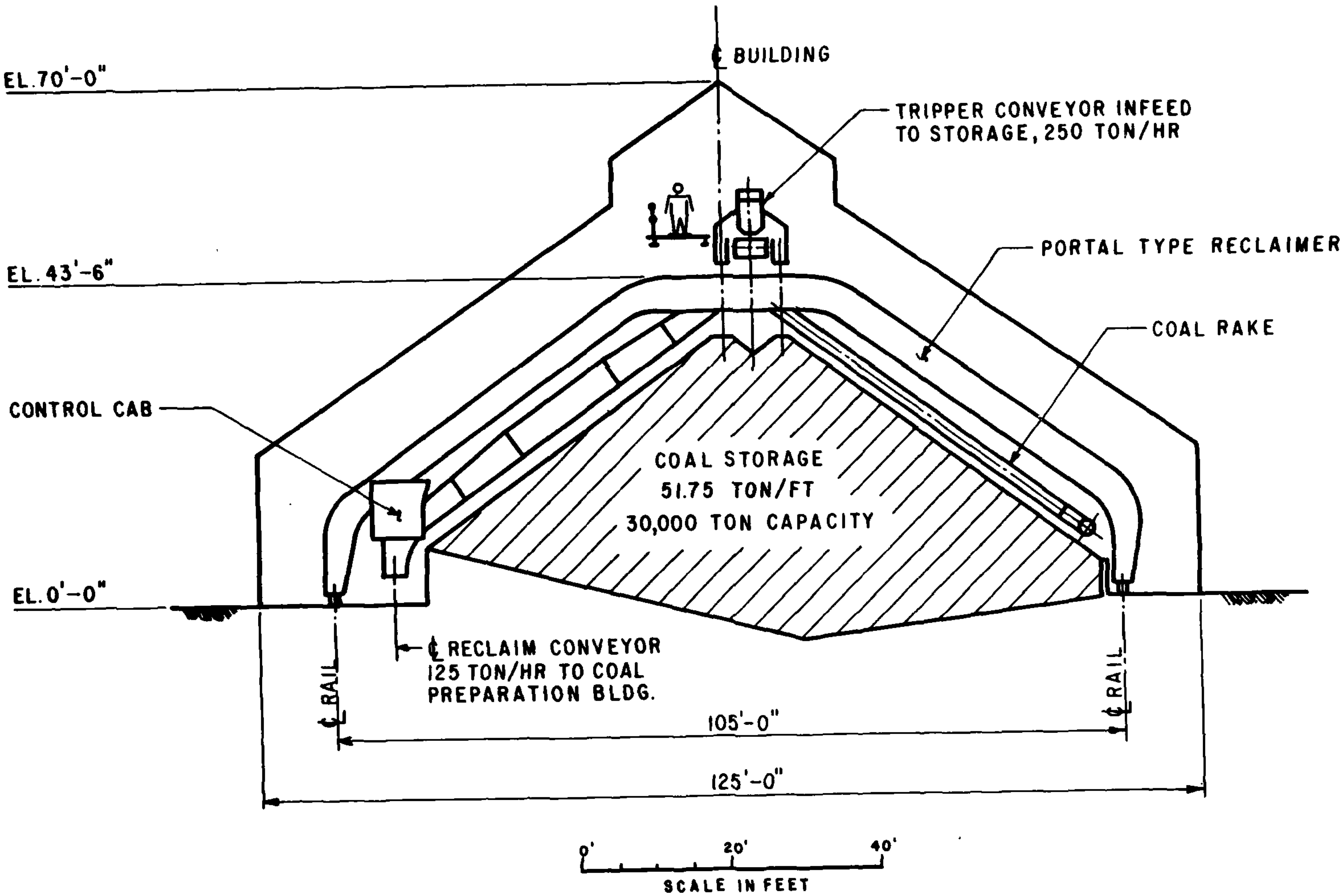
- Tripper conveyor to distribute coal along a 650 foot long pile. Tripper will be fitted with adjustable discharge spouts to avoid dusting.

- Portal type rake reclaimer to gather coal from the pile to a belt conveyor which leads to the coal preparation building. Reclaim is at the rate of 125 tons/hour.

All of the above equipment is standard for industrial plants that fire coal. The portal type reclaimer is more evident in European plants and is shown in Exhibit 7-3. It permits automated reclaim at high rates from an enclosed pile. It would be possible to store coal within a walled enclosure provided that the runoff was collected and treated. However, the proximity of non-industrial use neighbors precludes open storage. None of the conveying equipment is exposed, while the unloading operation is further protected from emitting dust by provision of a dust control and filtration system.

A weigh scale will be installed on the belt leaving the skip hoists so that inventory control is possible. This will also act as a check against bills of lading from the coal supplier.

COAL STORAGE BUILDING
SECTION LOOKING WEST



POPE, EVANS AND ROBBINS

EXHIBIT 7-3

Coal will also be weighed as it is fed into the process as a check on unit efficiency.

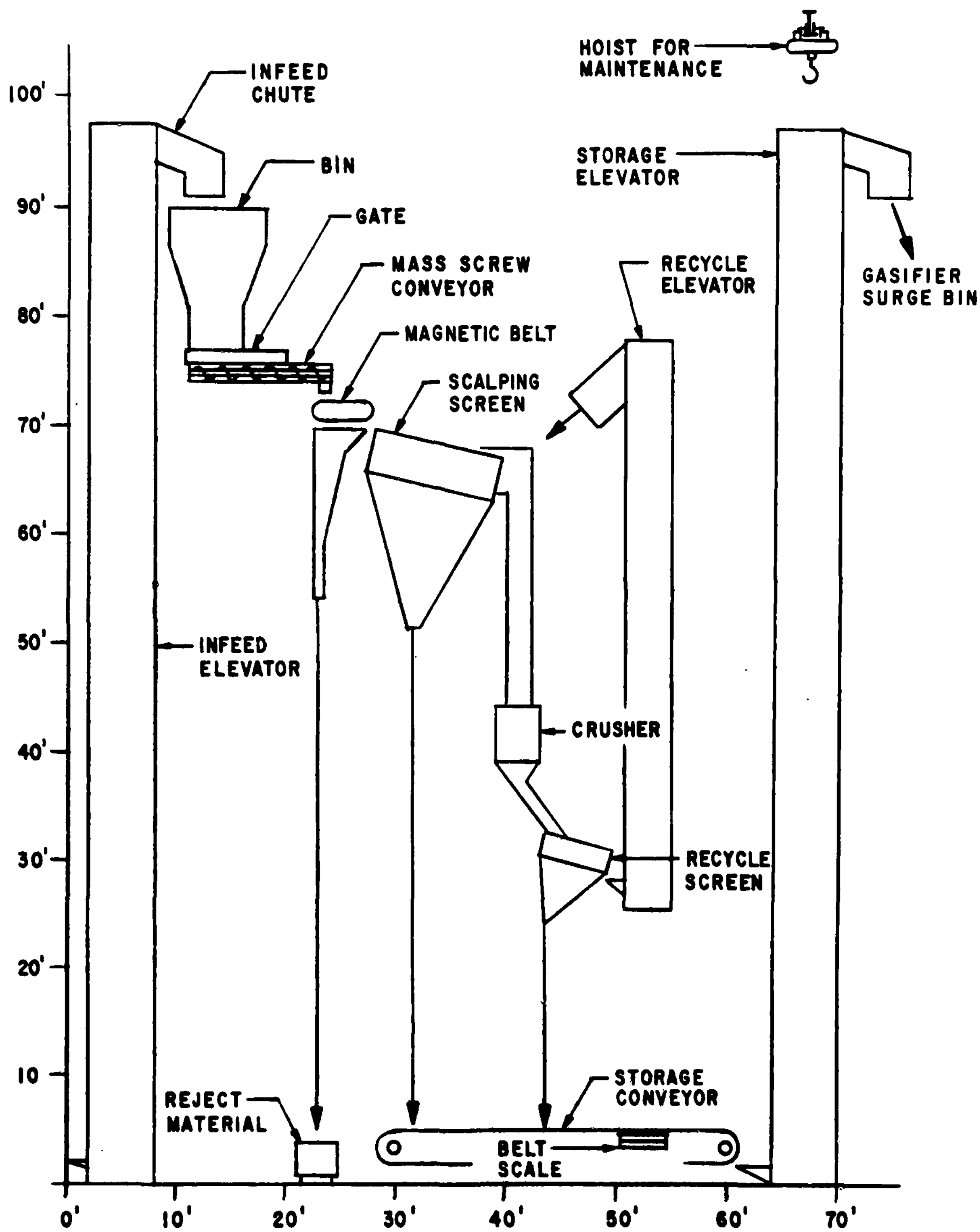
Coal from storage will be prepared in accordance with the needs of the particular gasifier supplier. Coal delivered to the preparation area will be elevated to the highest point for a downward travel through a scalping screen. The undersized coal will be collected and placed on a belt for delivery to the gasifier. Oversize coal will pass through a crusher to suit the size requirements. In order to avoid extremely fine coal, a screen at the crusher discharge will pass sized coal to the collecting belt, while oversize coal will be returned for recrushing. Duplication of equipment will assure reliability. See Exhibits 7-4 and 7-5. No attempt will be made to beneficiate coal, assuming that it is a washed product. Considering the quantities involved, it will not be economical to order run-of-mine coal unless a coal breaker is installed. This sizing device rejects a percentage of the rock but represents a major capital investment and high maintenance costs. Selection of this equipment is left for final design considerations.

A small surge bin would be included in the discharge of the collecting conveyor so that the gasifier could call upon a steady stream of fuel with the ability to reduce flow without uncontrolled stoppage of the conveyor links preceding the sizing equipment. Coal from the surge bin would be elevated to the entry point air locks and be delivered by enmasse type enclosed conveyors.

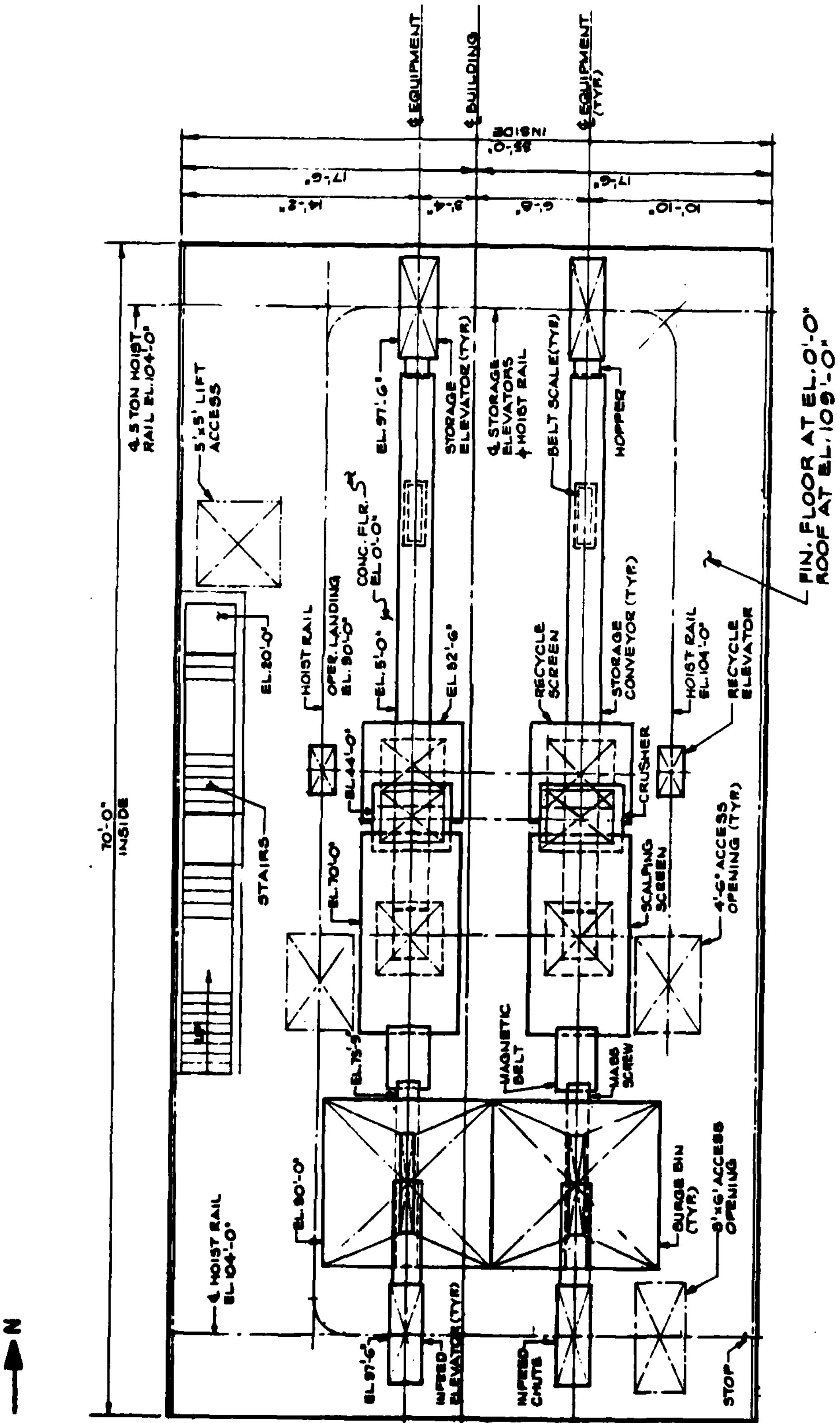
Other elements required on the site shown on Exhibit 7-2 include:

- Gasifier plant with its attendant clean-up process.
- Oxygen plant including storage of product.
- Combined cycle plant.

SCHEMATIC OF COAL PREPARATION BUILDING



COAL PREPARATION BUILDING PLAN



POPE, EVANS AND ROBBINS

EXHIBIT 7-5

- Reject product storage. This includes coal ash, tar produced by some gasification processes, washdown water for recycle and gases from the oxygen production process.

The gasification and cleanup area plant would be set adjacent to the coal preparation area with sufficient distance to provide a fire break and access for emergency vehicles. Access is also required for the collected tar as this is saleable in some systems and used in combustion in others. The degree of reject from the cleaned gas stream also depends upon the cleanup system that is used. Sulfuric acid and elemental sulfur are two common products. They may be saleable but temporary storage is required on-site to acquire sufficient quantity for disposal.

The oxygen preparation plant uses ambient air as a plant feed and produces nitrogen as a by-product during the process. The analysis takes no account of the marketing of nitrogen and no storage is planned, the gas being rejected to atmosphere. Oxygen is piped to the gasification plant over a separate trestle than the product gas which is piped to the combined cycle plant.

The combined cycle plant houses the energy conversion elements in the overall process. The structure will house:

- Air inlet filter silencer.
- Air compressor.
- Gas turbine, shaft connected to an electric generator.
- Switchgear and controls for frequency and voltage.
- Waste heat boiler with economizer and superheater.

- High pressure steam turbine directly connected to an electric generator.

- Outlet gas stack and silencer.

- Water conditioning equipment including feedwater receiver, transfer pumps, deaerating feedwater heater, and chemical feed tanks and pumps for feedwater and boiler application.

- Miscellaneous plant equipment including air compressor for plant cleanup and instrumentation, combustion control system, electrical synchronizing panels, alarm condition status board and supervisory information system.

The combined cycle plant is a vertically oriented building containing the air compressor-gas turbine-electric generator unit on the ground floor. Parallel to this will be the steam turbine and its generator. This provides for isolated structural support of these high rotational speed units. The waste heat boiler will be suspended from high roof steel to permit thermal growth downward, as is common for most high pressure steam generators. The deaerator will be midway up the building to assure sufficient head on the boiler feed pumps and the feedwater receiver will be set adjacent to the deaerator. Electric panels will be near the generators with a control room set at one end of the building overlooking the turbines. A man-lift will provide ready access within the building, with an outside hoist for equipment removal and heavy maintenance material.

All piping trestles, galleries and other connections between buildings will be maintained at a height of 20 feet above grade to permit unhindered movement of emergency vehicles in the area.

Access to ash, resulting from the gasification process, is also required. Ash will be produced in proportion to that delivered with the coal. For a nominal 10 percent ash content, ash will be produced at the rate of 100 ton/day and occupy a volume of approximately 8000 cubic feet. If allowance is made for a 5 day accumulation, the ash silos may be sized as two steel tanks, each 25 foot diameter by 42 foot on the shell with conical bottoms and cone dischargers. Each would be fitted with an ash conditioner to reduce escape of dust. The overall height would be 87 feet plus 6 feet for a vent filter.

The waste water recycle tank will hold 100,000 gallons and be sized as a cylindrical tank with conical roof, supported on grade. It would be sized as 30 foot diameter by 25 foot high. Discharge of sludge from this tank would require treatment before acceptance by the sanitary sewer system.

Coal storage volume for the conventional plant would be less than for the coal gasification plant. Silo storage would provide the environmental protection required and make coal available on demand. A 225,000 lb/hr boiler would need four 50-foot diameter by 80 feet high silos, while the three 200,000 lb/hr boilers would use four 60-foot diameter by 90 foot high silos to provide 30 days of storage. The 30-day storage requirement might be lessened, considering the ability of these boilers to fire No. 6 fuel oil as well as coal.

Ash storage for the conventional plant would be similar to the coal gasification plant requirements, however, there are other reject products: sludge for a wet scrubber system and spent sorbent for a dry scrubber or a fluid bed combustion process. Enclosed storage for these products would be required prior to their disposal offsite.

8.0 ECONOMIC ANALYSES AND RECOMMENDATIONS

In this section economic evaluation is made of the candidates previously identified as technically feasible and suitable for consideration as cogeneration additions for SPNC. This evaluation is performed on the basis of analyses of life cycle costs and life cycle energy requirements.

In Section 2, the following six gasifiers were selected as representative of those commercially available:

Fixed bed gasifiers	Lurgi, dry ash Wellman-Galusha Woodall-Duckham
Fluidized bed gasifiers	Winkler
Entrained bed gasifiers	Koppers-Totzek Texaco

Consider first some preliminary remarks. The Winkler gasifier is available only in sizes requiring 700 to 1,000 tons/day of coal fuel. The coal fuel input for a coal gasification/combined cycle (cg/cc) plant producing 50 MW of electric power ranges from 700 to 800 tons/day, depending on the gasifier used. Therefore, use of the Winkler gasifier would require installing two gasifiers to provide any reasonable reliability; this would be clearly uneconomic in this size application.

The Wellman-Galusha gasifier has had experience only in an air-blown mode thus providing only low Btu gas productive capability. For use in the gas turbine stage of this project, medium Btu gas is an economic requirement.

For these reasons, the economic evaluation of the coal gasification has been narrowed to four gasifiers: two fixed beds - Lurgi and Woodall-Duckham and two entrained beds - Koppers-Totzek and Texaco. The analysis of Section 6.0 indicated that a 50 MW cogeneration scheme was on optimal size. Accordingly, the

performance summary for the cg/cc plant producing 50 MW electric power using each of these candidate gasifiers and derived from the data in Section 3.0, is shown in Table 8-1.

For comparison to a cg/cc plant, two conventional coal-fired cogeneration plant candidates are considered. The first system provides for the addition of one 225,000 lb/hr coal-fired boiler and the other, the addition of three 200,000 lb/hr coal-fired boilers.

The economic evaluation is based on a 25 year plant life and assumes plant start up in 1988. The base case used for comparison and identification of life cycle cost savings is:

- Plant P-1 continuing to be oil-fired as presently operated, with total purchase of electricity.

- Plant P-1 to be coal and oil-fired after four existing boilers are converted to coal-firing under a recent project (P-985); we note that it has recently been deprogrammed. Electricity continues to be purchased.

Table 8-2 contains the 1981 and 1988 capital and operating costs which comprise each of the above base cases.

The analyses are made using the escalation rates and methodology prescribed by NAVFAC and DOE in:

- LANTNAVFACENGCOM, Instruction for Preparation of Economic Analyses, 407:ARM, 19 March 1980. (NAVFAC) (see Reference 1).

- Federal Register 45 209, October 27, 1980 (DOE) (see Reference 2).

TABLE 8-1

COAL GASIFICATION/COMBINED CYCLE PLANT

	Lurgi	Woodall-Duckham	Koppers-Totzek	Texaco
Energy Input (10^6 Btu/hr)	789	842	772	770
Annual Coal Use (Tons)	281,000	300,000	326,500	274,300
Net Electric Generated (MW)	50	50	50	50
Net Export Steam Flow (lb/hr)	194,000	241,000	275,000	294,000
Thermal-To-Electric Ratio	1.37	1.70	1.98	2.07
Overall Thermal Efficiency	0.51	0.54	0.64	0.68

TABLE 8-2

BASE CASES
CAPITAL AND OPERATING COSTS[1]

	Oil-Fired Present Plant		Coal-Fired Converted Plant	
	$ 1981	$ 1988	$ 1981	$ 1988
Capital Cost	-	-	25,350[2]	40,660[2]
Operating Costs:				
Operations and Maintenance	5,790	8,480	6,760	9,560
Oil Fuel	17,750	44,410	-	-
Coal Fuel	3,640	7,080	10,250	19,980
Electricity	31,600	74,330	31,600	74,330

1. All dollars are in 1,000.
2. Derived from P-985 data.

8.1 NAVFAC and DOE: Comparison of Escalation Rates and Methodology

There are substantial differences between the NAVFAC and DOE approach to economic analyses with regard both to the prescribed escalation rates and the methodology to be followed. For the NAVFAC approach:

- Analyses are made on the basis of project-start dollars, 1988 in this case.
- Current capital, operating and fuel costs are escalated to 1988 on the basis of prescribed short-term escalation rates which include general inflation.
- Long term differential escalation rates (not including general inflation) are applied to the 1988 fuel costs from project start to project end.
- Discounting is performed at 10% per year.

For DOE:

- Analyses are made on the basis of 1981 dollars.
- Short term escalation rates are not to be applied to current capital, operating and fuel costs. Differential rates for fuel are provided for the period 1981 to 1988.
- Long term differential rates are applied to the 1988 fuel costs to project end.
- Discounting is performed at 7% per year.
- Economic cost effectiveness is measured against 90% of the actual investment. This 10% investment credit is a proxy for externality adjustments and represents the

DOE evaluation of the benefit of fossil fuel conserving investments.

The prescribed escalation rates for NAVFAC and DOE are shown in Table 8-3. The discounted uniform present worth factors for the 25 year project life compare as follows:

	NAVFAC	DOE
Residual Oil	14.78	18.15
Coal	20.05	13.62
Electric	18.05	11.16

The most fundamental difference in approach is the use of 1988 dollars in one case and 1981 dollars in the other; this makes comparison between the two approaches impossible and prevents evaluation of the impact of the differing escalation scenarios prescribed. To allow such evaluation, we have developed self-consistent short escalation rates for DOE. Assuming a general inflation of 8% per year, we have added that to the prescribed DOE short term escalation rates and then have performed analysis on the basis of 1988 dollars.

It should be noted that, in addition to the differences in approaches used, there is a difference in project evaluation criteria applied by NAVFAC and by DOE. The important differences are:

- A project is considered economically effective by DOE if the discounted savings/investment ratio (SIR) is at least 1. NAVFAC generally requires SIR to be a substantial multiple of 1.

- DOE accepts a simple payback period "significantly less" than the projected life. NAVFAC measures the discounted payback period and generally requires it to be a small fraction of the project life.

TABLE 8-3

COMPARISON OF ESCALATION RATES

NAVFAC

Discount Rate 10%

	Short Term Escalation Rates	Long Term Differential Escalation Rates*
Operations and Maintenance	5.6%	0.0%
Oil	14.0%	8.0%
Coal	10.0%	5.0%
Electric	14.0%	7.0%

US DOE

Region 3, including Virginia

Discount Rate 7%

	Long Term Differential Escalation Rates* - U.S. Average		
	1980 - 1985	1985 - 1990	1990 and Beyond
Operations and Maintenance	0.0%	0.0%	0.0%
Residual Oil	7.53%	2.58%	4.28%
Coal	9.63%	1.97%	1.12%
Electric	-0.01%	1.19%	-0.50%

*Differential escalation rates are defined to be those above (or below) inflation.

As a measure of providing another project evaluation criteria we have developed the return on equity for each candidate; this is a measure of great importance to the private sector investor.

8.2 Evaluation of the Gasification Process

Prior to carrying out the detailed life cycle costs for the various complete systems, it seems appropriate to first assess the validity of gasification alone. Here we wish to evaluate whether, given current coal and gasifier costs, such a process is indeed economic compared to purchase of natural gas.

We do so for the two discount rates suggested by NAVFAC and DOE and for a range of gasifier costs (see Section 8.3 below for details of these values); results are shown in Exhibit 8-1. Inspection suggests that natural gas costs would have to become considerably lower to degrade the economic benefit of coal gasification.

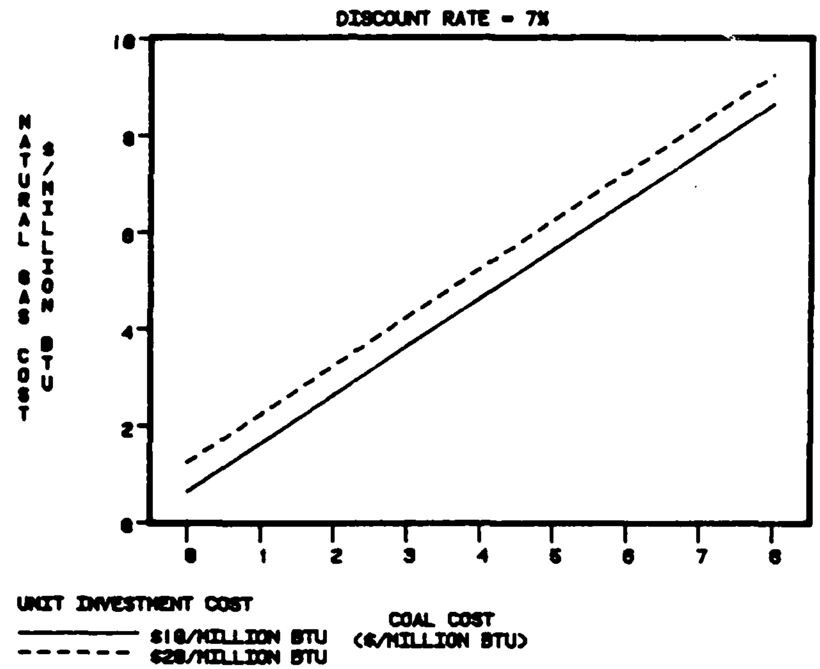
8.3 Screening the CG/CC Plant Candidates

A screening of the cg/cc plant candidates with the four remaining representative gasifiers was carried out on the basis of NAVFAC escalation using the present plant on oil as the base case. This will provide a single gasifier system to assess in some detail.

Current capital costs and estimated operating costs were provided by the suppliers only for the Woodall-Duckham gasifier. For the other gasifiers, we had to rely on the literature (see References 3 to 8) for development of such cost information. Capital costs for the Koppers-Totzek gasifier could not be developed and for the purposes of comparison we assumed the capital cost to be the same as for Texaco, the other entrained bed gasifier. Current cost information was received from the suppliers for the combined cycle portion of the plant.

EFFECT OF COAL GASIFICATION COSTS

25 YEAR LIFE CYCLE EVALUATION



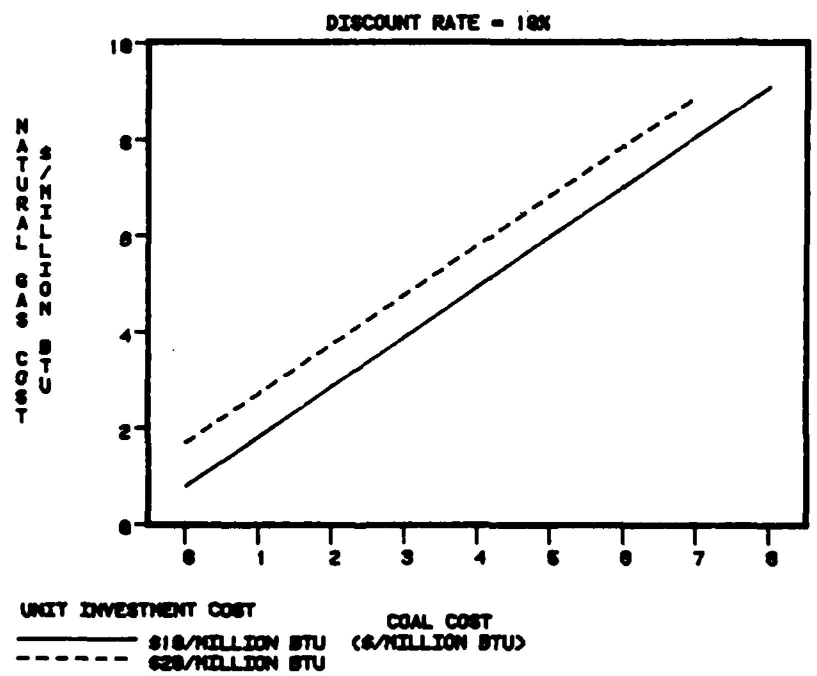


EXHIBIT 8-1

Capital costs are shown in Tables 8-4, 8-5, 8-6 and 8-7. In addition to the actual cost line items, we have indicated unit costs, as well, in terms of both gasifier output (10^6 Btu/yr) and combined cycle output (kW). From these we may observe tht coal handling-through-gas-production yields a cost of roughly $\$14/10^6$ Btu/yr, without engineering, startup and contingency and independent of gasifier. Cycle costs are roughly $775/kW. It is also of interest to derive total unit costs. These range from $\$28/10^6$ Btu/yr to $\$31/10^6$ Btu/yr or $2870/kW to $3070/kW.

Operating costs for the four gasifiers are shown in Table 8-8. Again, units costs of operation are provided and clearly show the advantages of the cogeneration schemes.

The results of the life cycle cost analyses are shown in Table 8-9. It can be seen that the entrained bed gasifier cg/cc plants are more economically effective than those using fixed bed technology. Furthermore, while the differences between the Texaco and the Koppers-Totzek cases are very marginal, we need to select one for additional analysis; Texaco has been chosen.

8.4 Screening the Conventional Coal-Fired Cogeneration Plant Candidates

For each of the two conventional coal-fired cogeneration plant candidates, three possible boiler configurations were considered; stoker fed with flue gas desulfurization (Case I), pulverized with flue gas desulfurization (Case II), and fluidized bed combustion (Case III). Capital costs of each of the configurations are set forth in Tables 8-10 and 8-11. For each conventional plant candidate, the configuration using fluidized bed technology has lower first cost; therefore, it is this configurtion which we use for further comparison of the conventional candidates. Operating costs are shown in Table 8-12.

TABLE 8-4

COAL GASIFICATION/COMBINED CYCLE PLANT
CAPITAL COSTS

A. LURGI

Line Item	Cost ($000)	Unit Cost $/10^6 Btu/yr	Unit Cost $/kW	%
1. Coal Receiving and Storage	13,650	2.7		12.8
2. Coal Handling	5,090	1.0		4.8
3. Oxygen Plant	11,200	2.2		10.5
4. Gasification Plant	21,300	4.2		20.0
5. Process Effluent Water Treatment	6,330	1.2		5.9
6. Gas Cleanup and Sulfur Recovery	10,100	2.0		9.5
SUBTOTAL	67,670	13.3		
7. Combined Cycle	30,700		614	28.8
8. Utilities and Waste Disposal	2,000		40	1.9
9. Electrical Distribution	1,000		20	1.0
10. Site Work, Foundations, Support Facilities	5,070		101	4.8
SUBTOTAL	38,770		775	100.0
TOTAL	106,440			
Engineering (10%)	10,640			
	117,080			
Start-Up (2%)	2,340			
	119,420			
Contingency and SIOH (20%)	23,880			
GRAND TOTAL	143,300			

1. All dollars are in 1988.

TABLE 8-5

COAL GASIFICATION/COMBINED CYCLE PLANT
CAPITAL COSTS

B. WOODALL-DUCKHAM

Line Item	Cost ($000)	Unit Cost $/10 Btu/yr	Unit Cost $/kW	%
1. Coal Receiving and Storage	13,650	2.5		12.7
2. Coal Handling	5,090	0.9		4.7
3. Oxygen Plant	16,720	3.1		15.5
4. Gasification Plant	29,200	5.4		27.1
5. Process Effluent Water Treatment	4,200	0.8		3.9
6. Gas Cleanup and Sulfur Recovery	(Note 2)	-		-
SUBTOTAL	68,860	12.7		
7. Combined Cycle	30,700		614	28.5
8. Utilities and Waste Disposal	2,000		40	1.9
9. Electrical Distribution	1,000		20	.9
10. Site Work, Foundations, Support Facilities	5,130		101	4.8
SUBTOTAL	38,830		775	100.0
TOTAL	107,690			
Engineering (10%)	10,770			
	118,460			
Start-Up (2%)	2,370			
	120,830			
Contingency and SIOH (20%)	24,170			
GRAND TOTAL	145,000			

1. All dollars are in 1988.
2. Included in gasifier.

TABLE 8-6

COAL GASIFICATION/COMBINED CYCLE PLANT
CAPITAL COSTS

C. KOPPERS-TOTZEK

Line Item	Cost ($000)	Unit Cost $/10 Btu/yr	Unit Cost $/kW	%
1. Coal Receiving and Storage	13,650	2.6		12.0
2. Coal Handling	4,500	0.9		3.9
3. Oxygen Plant	23,600	4.5		20.7
4. Gasification Plant	20,500	3.9		18.0
5. Process Effluent Water Treatment	1,300	0.2		1.1
6. Gas Cleanup and Sulfur Recovery	11,400	2.2		10.0
SUBTOTAL	74,950	14.3		
7. Combined Cycle	30,700		614	26.9
8. Utilities and Waste Disposal	2,000		40	1.8
9. Electrical Distribution	1,000		20	0.9
10. Site Work, Foundations, Support Facilities	5,400		108	4.7
SUBTOTAL	39,100		782	100.0
TOTAL	114,050			
Engineering (10%)	11,405			
	125,455			
Start-Up (2%)	2,510			
	127,965			
Contingency and SIOH (20%)	25,635			
GRAND TOTAL	153,600			

1. All dollars are in 1988.

TABLE 8-7

COAL GASIFICATION/COMBINED CYCLE PLANT
CAPITAL COSTS

D. TEXACO

Line Item	Cost ($000)	Unit Cost $/10 Btu/yr	Unit Cost $/kW	%
1. Coal Receiving and Storage	13,650	2.8		12.0
2. Coal Handling	4,500	0.9		3.9
3. Oxygen Plant	23,600	4.8		20.7
4. Gasification Plant	20,500	4.1		18.0
5. Process Effluent Water Treatment	1,300	0.3		1.1
6. Gas Cleanup and Sulfur Recovery	11,400	2.3		10.0
SUBTOTAL	74,950	15.2		
7. Combined Cycle	30,700		614	26.9
8. Utilities and Waste Disposal	2,000		40	1.8
9. Electrical Distribution	1,000		20	0.9
10. Site Work, Foundations, Support Facilities	5,400		108	4.7
SUBTOTAL	39,100		782	100.0
TOTAL	114,050			
Engineering (10%)	11,405			
	125,455			
Start-Up (2%)	2,510			
	127,965			
Contingency and SIOH (20%)	25,635			
GRAND TOTAL	153,600			

1. All dollars are in 1988.

TABLE 8-8

COAL GASIFICATION/COMBINED CYCLE PLANT
OPERATING COSTS

	Lurgi	Woodall-Duckham	Koppers-Totzek	Texaco
Labor	5,680	5,240	5,240	5,530
Maintenance Material	2,400	1,760	1,900	1,900
Plant Startup Materials	1,000	760	1,000	1,000
Ash Removal	310	330	310	300
Catalyst and Chemicals	200	180	190	180
Car Movement	250	260	250	240
Sulfur Credit	(90)	(100)	(90)	(90)
Materials for Balance of Plant[2]	1,120	880	690	650
TOTAL O&M	10,870	9,310	9,490	9,710
Unit Costs:				
¢/kWh/yr	2.5	2.1	2.2	2.2
¢/lb of steam/yr	0.6	0.4	0.4	0.4

NOTES:

1. All dollars are in 1981 x 1000.
2. Based on continued operation of balance of existing plant on oil-firing; this will increase by 150 if balance of plant is converted to partial coal-firing.

TABLE 8-9

LIFE CYCLE COST ANALYSIS
COAL GASIFICATION/COMBINED CYCLE PLANT
VS. CONTINUED OPERATION OF OIL-FIRED PLANT
USING NAVFAC APPROACH

	Lurgi	Woodall-Duckham	Koppers-Totzek	Texaco
Investment	226,700	229,400	243,000	243,000
Total Life Cycle Savings	1,121,260	1,185,970	1,288,340	1,301,320
Savings Investment Ratio	4.95	5.17	5.30	5.36
Discounted Payback Period (Years)	4.48	4.26	4.07	4.06
Return on Equity (%)	36.7	38.2	39.6	39.7
ECR (10^6 Btu/$1000 Investment)	3.9	4.3	7.2	8.3

1. All dollars are in 1988 x 1000.

TABLE 8-10

CONVENTIONAL COAL-FIRED COGENERATION PLANTS
CAPITAL COSTS

A. One 225,000 lb/hr Boiler and One 5,000 kW Turbine

	Case I Note 2	Case II Note 3	Case III Note 4
1. Coal Receipt, Storage and Delivery	7,300	7,300	7,300
2. Interior Alternations	580	620	740
3. Boiler and Burners	8,400	11,170	13,000
4. Combustion Control	500	700	200
5. Turbine-Generator and Switchgear	2,200	2,200	2,200
6. Piping With Insulation	1,600	1,600	1,800
7. Stack Gas Cleanup	5,860	5,960	550
8. Electrical Work	700	730	880
9. Structural/Architectural	800	1,000	1,200
	27,940	31,280	27,870
Contractor Overhead and Profit (25%)	6,985	7,820	6,968
	34,925	39,100	34,838
S.I.O.H. (5.5%)	1,921	2,150	1,916
	36,846	41,250	36,754
Contingency (10.0%)	3,685	4,125	3,675
TOTAL	40,531	45,375	40,429

1. All dollars are in 1981 x 1000.
2. Stoker-fed with flue gas desulfurization.
3. Pulverized-coal-fed with flue gas desulfurization.
4. Fluidized bed combustion.

TABLE 8-11

CONVENTIONAL COAL-FIRED COGENERATION PLANTS
CAPITAL COSTS

B. Three 200,000 lb/hr Boilers and One 13,000 kW Turbine

	Case I Note 2	Case II Note 3	Case III Note 4
1. Coal Receipt, Storage and Delivery	7,300	7,300	7,300
2. Interior Alternations	2,020	2,140	2,500
3. Boiler and Burners	22,050	29,830	34,700
4. Combustion Control	1,500	2,100	600
5. Turbine-Generator and Swtichgear	4,850	4,850	4,850
6. Piping With Insulation	4,800	4,800	5,400
7. Stack Gas Cleanup	17,580	17,880	1,650
8. Electrical Work	1,760	1,850	1,850
9. Structural/Architectural	2,400	3,000	3,600
	64,260	73,750	62,450
Contractor Overhead and Profit (25%)	16,065	17,437	15,613
	80,325	91,187	78,063
S.I.O.H. (5.5%)	4,418	5,015	4,293
	84,743	96,202	83,356
Contingency (10.0%)	8,474	9,620	8,336
TOTAL	93,217	105,822	91,682

1. All dollars are in 1981 x 1000.
2. Stoker-fed with flue gas desulfurization.
3. Pulverized-coal-fed with flue gas desulfurization.
4. Fluidized bed combustion.

TABLE 8-12

CONVENTIONAL COAL-FIRED COGENERATION PLANT
OPERATING COSTS

		Case II With One 225,000 lb/hr Boiler	Case III With Three 220,000 lb/hr Boilers
1.	Labor	3,780	4,320
2.	Maintenance Material	220	480
3.	Limestone	500	960
4.	Ash Removal[2]	760	1,620
5.	Materials for Balance of Plant	1,510	270
6.	Total O&M	6,770	7,650

1. All dollars are in 1981 x 1000.
2. Includes spent limestone.

The results of the life cycle cost analyses using NAVFAC as screening are shown in Table 8-13. It can be seen that the addition of only one 225,000 lb/hr boiler is more economically effective; it is this candidate which will be compared to the Texaco cg/cc plant candidate.

8.5 Comparison of CG/CC Plant with Conventional Plant

Table 8-14 sets forth, using NAVFAC escalations, the details of the life cycle costs of the cg/cc plant and the conventional plant compared to the continued oil-firing of the current plant. Table 8-15 shows the similar life cycle costs summary using DOE escalations.

A summary of the life cycle analysis for both configurations compared to present operations on oil are set forth in Table 8-16 with results from both the NAVFAC and DOE approaches. This, in essence, provides the sensitivity band for fuel costs and escalation scenarios. Depending on the economics used, the results developed for the cg/cc plant alternative fall in the following ranges:

SIR	2.6 to 5.4
Discounted Payback Period	4.1 to 7.1 years
Return on Equity	19.3% to 39.7%

It can be seen that the addition of a coal gasificaton/combined cycle plant will result in a viable project and an economically attractive investment opportunity. To be noted also is that the ECRs (the energy/cost ratio - Mbtu per thousand dollars of investment) developed fall in the generally acceptable range.

It may also be seen that the corresponding conventional coal-fired cogeneration plant alternative results are measurably better than those developed for the cg/cc plant alternative.

TABLE 8-13

LIFE CYCLE COST ANALYSIS
CONVENTIONAL COAL-FIRED COGENERATION PLANT
VS. CONTINUED OPERATION OF OIL-FIRED PLANT
USING NAVFAC APPROACH

	One 225,000 lb/hr Boiler	Three 200,000 lb/hr Boilers
Investment	63,960	145,040
Total Life Cycle Savings	565,140	928,570
Savings Investment Ratio	8.84	6.40
Discounted Payback Period (Yrs)	2.44	3.46
Return on Equity (%)	65.6	46.4
ECR (10^6 Btu/$1000 Investment)	1.7	2.9

1. All dollars are in 1981 x 1000.

TABLE 8-14

SUMMARY OF LIFE CYCLE COSTS
CG/CC PLANT AND CONVENTIONAL PLANT
VS. CONTINUED OPERATION OF OIL-FIRED PLANT
USING NAVFAC APPROACH

		CG/CC Plant	Conventional Plant
1.	Capital Costs	$243,000	$63,960
2.	Operations and Maintenance Costs	54,800	13,630
3.	Oil Costs	(759,540)[2]	(462,580)[2]
4.	Coal Costs	332,260	127,850
5.	Electric Costs	(928,840)[2]	(462,580)[2]
6.	Life Cycle Savings	1,301,320	565,140

1. All dollars are in 1981 x 1000.
2. Represents savings.

TABLE 8-15

SUMMARY OF LIFE CYCLE COSTS
CG/CC PLANT AND CONVENTIONAL PLANT
VS. CONTINUED OPERATION OF OIL-FIRED PLANT
USING DOE APPROACH

	CG/CC Plant	Conventional Plant
1. Capital Costs	$263,200	$62,360
2. Operations and Maintenance Costs	78,310	19,580
3. Oil Costs	(642,190)[2]	(391,090)[2]
4. Coal Costs	388,230	33,320
5. Electric Costs	(444,530)[2]	(116,810)[2]
6. Life Cycle Savings	620,180	455,000

1. All dollars are in 1981 x 1000.
2. Represents savings.

TABLE 8-16

LIFE CYCLE COST ANALYSES
CG/CC PLANT AND CONVENTIONAL PLANT
VS. CONTINUED OPERATION OF OIL-FIRED PLANT

	CG/CC Plant		Conventional Plant	
	NAVFAC	DOE	NAVFAC	DOE
Investment	$ 243,000	$263,200	$ 63,950	$ 69,290
Total Life Cycle Savings	1,301,320	620,180	565,140	455,000
Savings Investment Ratio	5.36	2.62	8.85	7.30
Discounted Payback Period (Years)	4.06	7.2	2.44	2.39
Return on Equity (%)	39.7	19.3	65.6	49.1
Energy/Cost Ratio (10^6 Btu/$1000 Invested)	8.3	7.6	1.7	1.6

1. All dollars are in 1988 x 1000.

The energy/cost ratios obtained, however, are appreciably smaller than those achieved in the cg/cc plant alternative.

Table 8-17 sets forth the results of the life cycle cost analyses of the two configurations compared to a partial conversion to coal-firing at boilers in Plant P-1 (Project P-985). The results developed are not as attractive since, in this case, displacement of base case fuel is primarily of coal rather than of oil. As has been noted already, this project has recently been deprogrammed; it is not clear when and whether the project will be reprogrammed.

8.6 National Applicability

Inquiry was made as to the national applicability of cg/cc plants in replacing industrial steam generation facilities presently firing oil. Some measure of the number of industries with appropriate characteristics for such plant improvement and the amount of energy cosumed therein has been the subject of previous investigations (see e.g. References 4, 8 and 9). From this, industries which have the appropriate thermal-to-electric load ratio and can accomodate cg/cc plants within the nationally limited 80 MW cogeneration limit were identified. Energy consumption and the potential for economically viable substitution of existing oil usage by coal usage are summarized in Table 8-18. There exists a potential of decreasing present oil usage by 260 million barrels per year by the introduction of coal gasification into these industries.

8.7 Third Party Ownership

Nothwithstanding the excellent results for this project, and the fact that there appears to be considerable and widespread applicability, the very large first cost of the cg/cc would make it difficult for NAVFAC to appropriate funds for the implementation. However, it is just these feasible results

TABLE 8-17

LIFE CYCLE COST ANALYSES
CG/CC PLANT AND CONVENTIONAL PLANT
VS. PARTIAL COAL CONVERSION OF OIL-FIRED PLANT

	CG/CC Plant		Conventional Plant	
	NAVFAC	DOE	NAVFAC	DOE
Investment	$243,000	$263,200	$ 63,960	$69,290
Total Life Cycle Savings	712,840	184,380	198,480	59,950
Savings Investment Ratio	2.93	0.78	3.10	0.98
Discounted Payback Period (Years)	7.29	-	6.79	29.4
Return on Equity (%)	24.6	-	25.8	6.5
Energy/Cost Ratio (10^6 Btu/$\$10^3$ Invested)	8.3	7.7	1.7	1.6

1. All dollars are in 1988 x 1000.

TABLE 8-18

POTENTIAL ENERGY SAVINGS
FOR
COAL GASIFICATION/COMBINED CYCLE COGENERATION CYCLES
AT
SELECTED CANDIDATE INDUSTRIES

Industry	Process Energy Requirements (10^6 BOE/yr)	Average Electric Demand (kW) Per 10^6 Units	Units (10^6)	Potential Oil Savings (10^6 BOE/yr)	Potential Net Energy Savings (10^6 BOE/yr)
Newsprint	21.1	137	4.1 tons	15.8	3.8
Writing Paper	31.4	101	3.6 tons	10.3	2.5
Corrugated Paper	154.7	68	18.5 tons	35.5	8.5
Boxboard	47.1	89	4.9 tons	12.3	2.9
Petroleum	195.1	0.3	6143 runs	59.4	14.3
Steel Mill	789.0	26	155 tons	113.2	27.2
Gray Iron Foundry	29.9	58	9.8 tons	15.8	3.8
TOTAL	1268.3			262.3	63.0

1. Electric requirements derived from source energy at 11,600 Btu/kWh.
2. Derived from coal gasification/combined cycle performance: energy saved from utility = 11,600 Btu/kWh, energy saved from thermal recovery = 8600 Btu/kWh.
3. Coal requirements for system = 15,400 Btu/kWh.

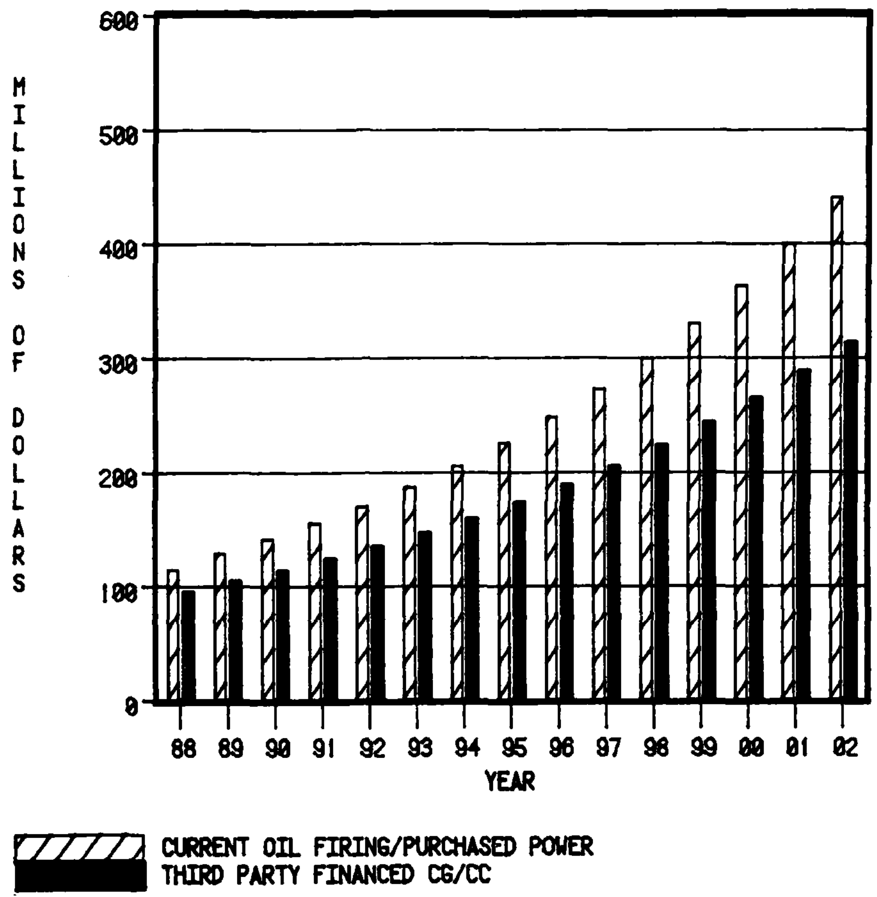
Note that BOE = barrels of oil equivalent.

which suggest the potential for active private sector investment and third-party financing of the project at SPNC.

We consider these options here but first distinguish the approach to the economics from the analyses already undertaken. All have been based on 100% equity, i.e. the public sector investor would supply the entire investment. The private sector investor, however, would structure his investment on the basis of leverage: the project capital cost being met partially by his investment (equity) and partially by bank loan (debt). Further, analyses herein are based on a zero-income-tax approach. The private sector investor naturally has to pay Federal and State income taxes but would have available investment tax credits, energy tax credits and newly accelerated depreciation rate schedules, all enacted specifically to encourage investment in projects precisely like the cg/cc. On these new bases then, there are created investment opportunities which should prove attractive to the private sector. We note that the opportunities for third-party projects are being increasingly appreciated (see Reference 10).

A typical result from third-party financing is shown in Exhibit 8-2 and Table 8-19. There we see, for a private sector investor with a goal of an after-tax return on equity of 20%, the structuring of such an investment. In the table we observe the stream of funds required retire the debt and to achieve the 20% return. Further, we show the savings which would accrue to SPNC for such a scheme. For NAVFAC, this would mean that, without any public sector investment or congressional appropriation, there exists an opportunity for SPNC to achieve $860,000,000 of the total savings over a postulated 15 year lease life. It should be noted that Table 8-19 is based on use of DOE escalation rates; use of private sector escalation rates would develop results even more attractive to the putative private investor and to SPNC (see Reference 11).

INTEGRATED COAL GASIFICATION/COMBINED CYCLE
COMPARISON OF ANNUAL OPERATING COSTS FOR SPNC



BASIS OF ANALYSIS (1981 DOLLARS)

TOTAL CONSTRUCTION COST	$154,000,000
CONSTRUCTION INTEREST	$15,400,000
INVESTMENT & ENERGY TAX CREDIT	$23,000,000
DEBT	$118,580,000
EQUITY	$50,820,000
DEBT LIFE (YEARS)	15
INTEREST ON DEBT (%)	17.0
TAX RATE (%)	46.0
RETURN ON EQUITY (%)	20.0

EXHIBIT 8-2

TABLE 8-19
FINANCIAL ANALYSIS PROGRAM
SEWELLS POINT NAVAL COMPLEX
THIRD PARTY FINANCING

YEAR	REVENUE	INTEREST	PRINCIPAL	DEBT SERVICE	DEPREC-IATION	PRE-TAX CASH FLOW	TAXABLE INCOME	TAXES	ETC+ ITC	AFTER-TAX CASH FLOW
0	0	0	0	0	0	0	0	0	39,480	39,480
1	37,690	34,453	3,612	38,065	44,542	-375	-41,304	-19,000	0	18,625
2	37,690	33,839	4,226	38,065	40,830	-375	-36,979	-17,010	0	16,635
3	37,690	33,120	4,944	38,065	37,118	-375	-32,548	-14,972	0	14,597
4	37,690	32,280	5,785	38,065	33,406	-375	-27,996	-12,878	0	12,503
5	37,690	31,296	6,768	38,065	29,694	-375	-23,301	-10,718	0	10,344
6	37,690	30,146	7,919	38,065	25,983	-375	-18,438	-8,482	0	8,107
7	37,690	28,800	9,265	38,065	22,271	-375	-13,380	-6,155	0	5,780
8	37,690	27,225	10,840	38,065	18,559	-375	-8,094	-3,723	0	3,348
9	37,690	25,382	12,683	38,065	14,847	-375	-2,539	-1,168	0	793
10	37,690	23,226	14,839	38,065	11,135	-375	3,329	1,531	0	-1,906
11	37,690	20,703	17,362	38,065	7,424	-375	9,563	4,399	0	-4,774
12	37,690	17,752	20,313	38,065	3,712	-375	16,227	7,464	0	-7,839
13	37,690	14,298	23,767	38,065	0	-375	23,392	10,760	0	-11,135
14	37,690	10,258	27,807	38,065	0	-375	27,432	12,619	0	-12,994
15	37,690	5,531	32,534	38,065	0	25,945	58,479	21,373	0	4,572
TOTAL	565,350	368,308	202,664	570,972	289,520	20,698	-66,158	-35,960	39,480	56,658

INPUT FOR 1988 DOLLARS	
TOTAL PROJECT COST	$289,520,000
COST OF EQUIPMENT	$263,200,000
FAIR MARKET VALUE	$26,320,000
CONSTRUCTION INTEREST	$26,320,000
INVESTMENT & ENERGY TAX CREDIT	$39,480,000
DEBT	$202,664,000
EQUITY	$86,856,000
DEBT LIFE	15
INTEREST ON DEBT	0.1700
TAX RATE	0.4600
RETURN ON EQUITY	0.2001

TABLE 8-19 (Cont'd)

NAVFAC CASH FLOW ANALYSIS
SEWELLS POINT NAVAL COMPLEX

YEAR	LEASE COST	COAL COST	O&M COST	ELECTRIC SAVINGS	OIL SAVINGS	NAVFAC SAVINGS
1	-37,690	-29,360	-6,720	38,740	36,470	1,440
2	-37,690	-32,038	-7,258	41,646	40,949	5,609
3	-37,690	-34,959	-7,838	44,769	45,977	10,258
4	-37,690	-38,148	-8,465	48,127	51,623	15,447
5	-37,690	-41,627	-9,142	51,736	57,962	21,239
6	-37,690	-45,423	-9,874	55,616	65,080	27,709
7	-37,690	-49,566	-10,664	59,788	73,0[illegible]2	34,940
8	-37,690	-54,086	-11,517	64,272	82,045	43,024
9	-37,690	-59,019	-12,438	69,092	92,120	52,065
10	-37,690	-64,401	-13,433	74,274	103,433	62,182
11	-37,690	-70,275	-14,508	79,844	116,134	73,506
12	-37,690	-76,684	-15,669	85,833	130,395	86,186
13	-37,690	-83,677	-16,922	92,270	146,408	100,389
14	-37,690	-91,309	-18,276	99,190	164,387	116,303
15	-37,690	-99,636	-19,738	106,630	184,574	134,139
TOTAL	-565,350	-870,208	-182,462	1,011,825	1,390,628	784,433

INFLATION RATE 0.08
DIFFERENTIAL ESCALATION RATES:

COAL	0.0112
ELECTRIC	-0.005
O&M	0
OIL	0.0428

8.8 Recommendations

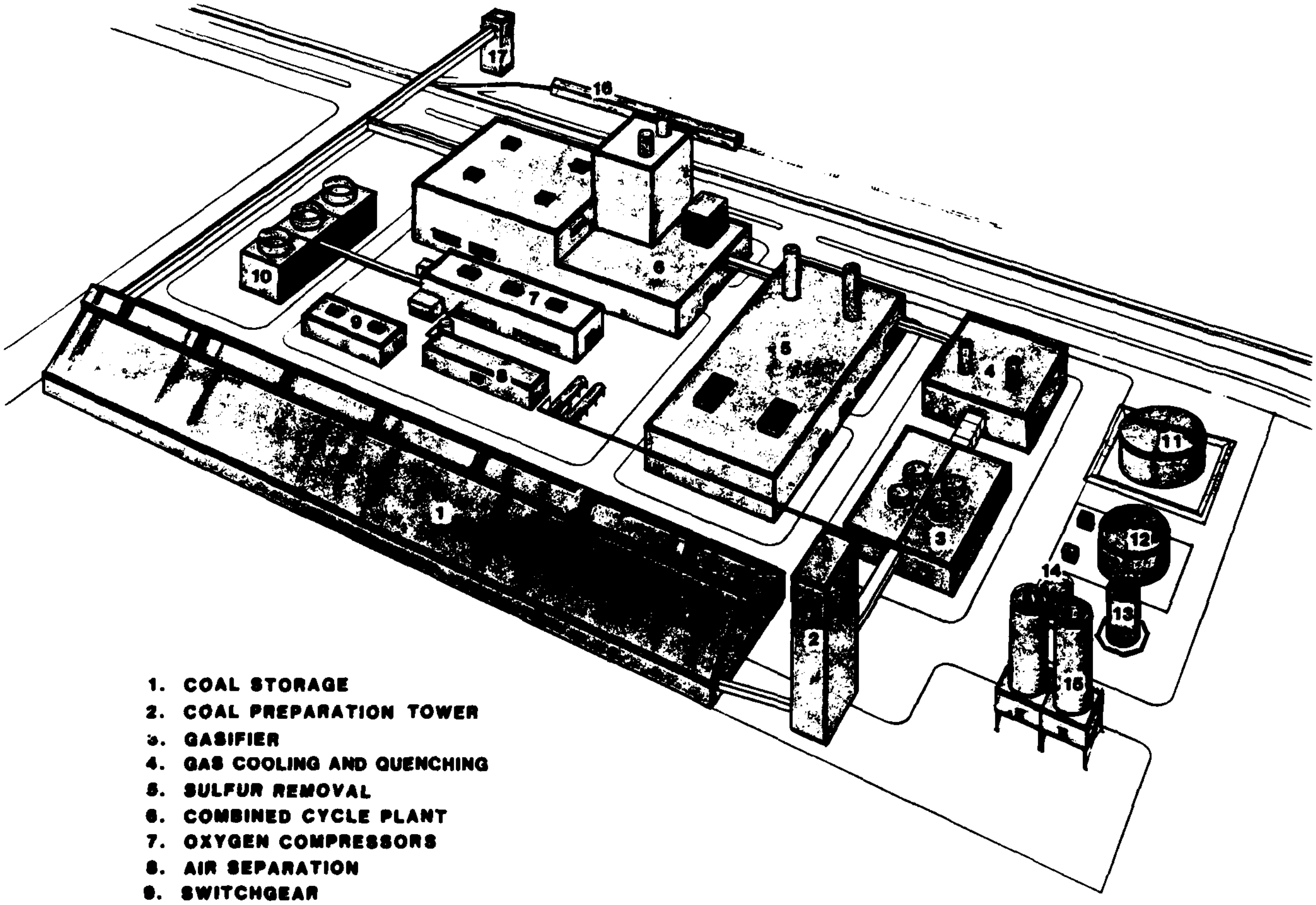
Current Administration concerns are "to establish sound policies that will encourage both the private and public sectors to produce and use energy resources wisely and efficiently". Further, there will be encouragement of an "increasing shift to coal-fired plants to replace more expensive oil" and "heavy reliance on private sector investment initiative in the synthetic fuel program". Clearly, this Energy Showcase project fulfills these goals.

Based on these results, the recommendations of this study are that:

- cogeneration at SPNC be pursued immediately,
- coal gasification/combined cycle be the technology employed at SPNC,
- the private sector be actively solicited for third-party financing, design and construction.

A typical plant layout of the cg/cc herein recommended for implementation at SPNC is shown in Exhibit 8-3.

COAL GASIFICATION COMBINED CYCLE COGENERATION PLANT



1. COAL STORAGE
2. COAL PREPARATION TOWER
3. GASIFIER
4. GAS COOLING AND QUENCHING
5. SULFUR REMOVAL
6. COMBINED CYCLE PLANT
7. OXYGEN COMPRESSORS
8. AIR SEPARATION
9. SWITCHGEAR
10. OXYGEN PLANT COOLING TOWER
11. TAR STORAGE (IF REQUIRED)
12. WASTE WATER TREATMENT
13. FIRE RESERVE
14. RECYCLE WATER
15. ASH SILOS
16. COAL TRAIN UNLOADING
17. COAL CONVEYOR TOWER

EXHIBIT 8-3

8.9 REFERENCES

8.1 LANTNAVFACENGCOM, Instruction for Preparation of Economic Analyses, 407:ARM, 19 March 1980.

8.2 Federal Register 45 209, October 27, 1980; see Ruegg, R.T., "Life Cycle Costing Manual for the Federal Energy management Programs", NBS Handbook 135, December 1980.

8.3 Gluckman, M.J. et al "Performance and Cost Characteristics of Combined Cycles Integrated with Second Generation Gasification Systems", ASME Paper 80-GT-106, see also Turbomachinery International, April 1981, pp 33-41.

8.4 "Industrial Cogeneration Optimization Program", prepared by TRW, Inc., Report DOE/CS/4300, September 1979.

8.5 "Economic Studies of Coal Gasification Combined Cycle Systems for Electric Power Generation", EPRI AF-642, Fluor Engineers and Constructors, January 1978.

8.6 "Preliminary Design Study for an Integrated Gas Gasification Combined Cycle Power Plant" prepared by Ralph M. Parsons Company, EPRI AF-880, August 1978.

8.7 "Coal Gasification Study" report prepared for NAVFAC Civil Engineering Laboratory, Port Hueneme, California, by Bechtel Corporation, April 1977.

8.8 "Cogeneration Technology Alternatives Study", United Technologies Corporation, several volumes, January 1981, NASA CR 159759, et. seq.

8.9 "Cogeneration Technology Alternatives Study", General Electric Company, several volumes, April 1980, NASA CR 159765, et. seq.

8.10 Baum, D. "Outside Investors Spark Growth of Cogeneration", Energy User News, Vol. 6, No. 29, Monday, July 20, 1981.

8.11 Katsidhe, V., et al, "Economic Considerations for Plant Investment", paper presented at the 16th IECEC, Atlanta, Georgia, August 1981.

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